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December 1978

EVALUATION OF PHYSICAL CHEMICAL TREATMENT AT ROSEMOUNT

by

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

A physical-chemical treatment plant designed to produce a very high quality effluent was placed in operation in 1973 in Rosemount, Minnesota, to treat the municipal wastewater. The U.S. Environmental Protection Agency and the Metropolitan Sewer Board agreed to jointly fund an evaluation of the new Rosemount facility. This publication reports the performance and cost data for this new technology.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

ABSTRACT

This research program was conducted to demonstrate the effectiveness of physical chemical treatment of raw municipal wastewater at the Rosemount Advanced Wastewater Treatment Plant and more specifically to evaluate: the performance of the system as a whole, the performance of the individual treatment processes, and the costs associated with operation and maintenance. During the two year demonstration period the facility treated an average flow of approximately 0.25 mgd by means of chemical clarification, filtration, carbon adsorption and ion exchange; both the activated carbon and ion exchange media were regenerated onsite.

The performance data are summarized according to the five process flow schemes used in addition to discussion of the individual treatment processes. Cost data are presented for each treatment process. These data are used to construct an estimate of operating and maintenance costs for a 10 mgd facility.

The operating and maintenance problems encountered during approximately four years of operation are described along with their solutions when they were determined. Design recommendations are presented.

This report was submitted in fulfillment of Grant No. S802666 by the Metropolitan Waste Control Commission under the partial sponsorship of the U.S. Environmental Protection Agency. This report covers the period from June 6, 1975 to August 13, 1977, and the work was completed as of January 12, 1978.

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LIST OF ABBREVIATIONS

cfm - cubic feet per minute

fps - feet per second

ft - feet

gm/cc - grams per cubic centimeter

gm. eq. wt - gram equivalent weight

gpd/sq ft - gallons per day per square foot

gph - gallons per hour

gpm - gallons per minute

hp - horsepower

hr - hour

in - inch

KWH - Kilowatt hour

lb - pound

lbs/cut ft - pounds per cubic foot - lbs/cu ft

ma - milliampere, direct current

me/gm - gram milliequivalent weights per gram of material

mgd - million gallons per day

mil gal - million gallons

min - minute

mm - millimeter

N - gram milliequivalent weights per liter of solution

psf - pounds per square foot

psi - pounds per square inch

sq ft - square foot

CONVERSION FACTORS

English Unit	multiply by	to yield Metric Unit
Cubic feet per minute	0.472	Liters per second
Feet	0.3048	Meters
Gallons	3.785	Liters
Gallons per day per square foot	0.352	Liters per day per square meter
Horse power	1.014	Horse power
Kilowatt - hours	860.5	Kilogram - calories
Million gallons	3785	Cubic meters
Pound	454	Grams
Pounds per cubic foot	16.02	Kilograms per cubic meter
Pounds per square foot	4.88	Kilograms per square meter
Pounds per square inch	0.0703	Kilograms per square centimeter
Square foot	0.0929	Square meters

I. INTRODUCTION

The Metropolitan Sewer Board, precursor of the current Metropolitan Waste Control Commission (MWCC) was formed in 1969. This agency is responsible for the operation of water pollution control facilities in the seven county metropolitan area in and around Minneapolis-St. Paul. In 1970 the agency acquired 33 wastewater treatment facilities along with 320 miles of interceptor sewer. Only four of the 33 plants produced treated effluents in compliance with applicable water quality standards at that time. Fifteen of the original plants have been closed to date and others have been upgraded. Three new treatment facilities including the Rosemount Advanced Wastewater Treatment Plant have been put into service since 1972.

The City of Rosemount is located 15 miles south of St. Paul and has a current population of approximately 4500. Prior to December 1973, the wastewater collected was treated at a high rate trickling filter plant and subsequently discharged to an effluent seepage pond. Both treatment plant and seepage pond were operating at their hydraulic capacity in 1970. Because there was no acceptable outlet for the plant effluent at the existing site, expansion of the existing facilities was not feasible. The decision was made to discharge the effluent from the new plant to the Mississippi River through Spring Lake. Spring Lake consists of the northwestern part of the impounded reach of the Mississippi River immediately upstream of Lock and Dam No. 2, located at Hastings, Minnesota. Proposals had been advanced to isolate the lake from the main river channel and develop its potential for water-based recreational use. The evaluation of the water quality standards and their applicability to receiving waters of limited dilution capacity led to the adoption of the effluent quality limits presented in TABLE 1 for design criteria.

TABLE 1. DESIGN CRITERIA

BOD	-	10 mg/l
Suspended solids	-	10 mg/l
Total phosphorus	-	1 mg/l
Ammonia nitrogen	-	1 mg/l
Total coliform organisms	-	100 MPN/100 ml

Several treatment alternatives were evaluated for a design flow of 0.6 MGD. Estimated capital and operating costs indicated that a physical-chemical treatment facility would be most cost efficient. (1) It was also recognized that the use of these new treatment processes would serve to fully evaluate the feasibility, economics and application of the individual unit processes for future treatment plant construction elsewhere.

The construction contract for the Rosemount Advanced Wastewater Treatment Plant (AWTP) was awarded December 15, 1971. In June of 1973 the U.S. Environmental Protection Agency and the Metropolitan Sewer Board agreed to jointly fund an evaluation and demonstration of the Rosemount facility. The specific objectives were to evaluate:

(a) performance of the system as a whole

(b) performance of individual unit operations and unit processes

(c) costs associated with operation and maintenance.

The major portion of the facility was put into service November 19, 1973. The zeolite regeneration and brine recovery system was not operational at that time however. The project agreement stipulated that the facility must be formally accepted by MWCC prior to the initiation of the demonstration program. The formal acceptance tests were concluded May 15, 1975 and the demonstration project was initiated June 6, 1975.

The purpose of this report is to summarize the information obtained by operating the Rosemount AWTP for the two year period June 6, 1975 through June 5, 1977. Additional data from the ammonia removal system are included for the period ending August 12, 1977 because they are significant in terms of evaluating that system.

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15. SUPPLEMENTARY NOTES

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16. ABSTRACT

This research program was conducted to demonstrate the effectiveness of physical chemical treatment of raw municipal wastewater at the Rosemount Advanced Wastewater Treatment Plant and more specifically to evaluate: the performance of the system as a whole, the performance of the individual treatment processes, and the costs associated with operation and maintenance. During the two year demonstration period the facility treated an average flow of approximately 0.25 mgd by means of chemical clarification, filtration, carbon adsorption and ion exchange; both the activated carbon and ion exchange media were regenerated onsite.

The performance data are summarized according to the five process flow schemes used in addition to discussion of the individual treatment processes. Cost data are presented for each treatment process. These data are used to construct an estimate of operating and maintenance costs for a 10 mgd facility.

The operating and maintenance problems encountered during approximately four years of operation are described along with their solutions when they were determined. Design recommendations are presented.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Sewage Treatment*, Chemical Removal*, Activated Carbon Treatment, Coagulation, Clarification, Filtration, Water Pollution	Physical-Chemical Treatment Ammonia Removal	13B

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II. SUMMARY AND CONCLUSIONS

A two year performance and cost evaluation program was conducted at the Rosemount AWTP. This physical chemical facility treated an average domestic wastewater flow of approximately 0.25 mgd by means of chemical clarification, filtration, carbon adsorption and ion exchange. Each process was evaluated in terms of both treatment characteristics and actual operating and maintenance costs. In addition total plant performance was evaluated during each of five flow schemes or modes of operation.

Although the average raw wastewater characteristics changed considerably during the treatment modes, the treatment efficiency in terms of percent removal of suspended solids, BOD and total phosphorus did not vary signifcantly and for the two year period removals averaged 99%, 93% and 94% respectively. The ammonia removal efficiency did vary substantially throughout the project as a result of modifications to the zeolite regeneration system. The process modifications were made to determine the effect of several parameters on the useful ammonia exchange capacity and associated operating costs.

The clarification process was aided by additions of lime and ferric chloride to provide for coagulation of finely divided solids and precipitation of soluble phosphorus. For a six month period the lime addition was regulated to maintain a pH of 10.5 in the clarifier. The lime dose was lowered during the following 18 month period to maintain a pH of 9.5. The ferric chloride dose was adjusted to maintain an effluent total phosphorus $\leq$ 1 mg/l. During the high pH operation the removal efficiencies averaged 94%, 81% and 95% for suspended solids, BOD and total phosphorus respectively. During this period the quality of the clarifier effluent approached that of a conventional secondary treatment facility. The performance degraded measureably when the pH was reduced to 9.5 and the respective removal efficiencies dropped to 88%, 75% and 88%. During both periods the clarifier exhibited a considerable capacity to accept shock loadings both in terms of flow and concentration and as a result significantly dampened possible suspended solids and BOD loading fluctuations on the downstream treatment units.

The dual media filters were located directly downstream of the clarifier (first stage) and carbon contractors (second stage). The suspended solids loading of the first stage filters averaged 0.47 psf per day and removals averaged 47%. The suspended solids loading of the second stage filters was considerably lower and removals averaged 30%.

During that period when the carbon columns were operated in the downflow mode the overall plant performance did not suffer when the second stage filters were removed from the flow scheme. However several attempts to operate the plant without the first stage filters failed because of rapid fouling of the second stage filters. Cleansing techniques using salt and/or chlorine along with normal air scour and backwash were developed.

The removal of organics across the carbon columns averaged 55% in terms of soluble COD. The removals were found to decrease as the cumulative loading (lb COD removed per lb carbon) increased. For the period during which the carbon columns were operated in the upflow mode the loading averaged 0.20 lb soluble COD removed per lb carbon which is considerably lower than several investigators have reported. The useful carbon loading rate increased to 0.4 lb per lb when the columns were operated in the downflow mode.

Daily air scour and backwash did not eliminate or control the production of hydrogen sulfide in the carbon beds and both odor and corrosion problems were encountered. It was found that hydrogen sulfide generation could be controlled when sodium nitrate was added to the column influent stream at a concentration of 5 mg/l NO_3-N and the columns were air scoured and backwashed daily.

Excessive carbon losses were encountered and appeared to be primarily related to the lack of screens at the backwash water discharge points and inadequate screening at the inlet to the screw conveyer serving the carbon regeneration furnace.

Carbon regenerations were conducted onsite with a four hearth gas fired furnace. Although carbon losses during regeneration were not documented it is estimated that they were minor compared to losses described above.

The ion exchange system used to remove NH_3-N was evaluated for nine operating schemes. It was found that the regenerant volume and pH were the most significant variables. The practical NH_3-N exchange capacity was found to fall in the range of 0.12 to 0.37 me/gm depending on regeneration procedure.

The actual unit cost of treatment was extremely high, $5.11 per 1000 gal, because the plant capacity and flow were low enough to preclude any economy of scale yet the plant complexity required full staffing. If the plant had operated at its design capacity during the two year period, the unit cost of treatment would have been reduced by approximately 50%. Although treatment facilities of this type may not be cost effective for small flows, operating costs for larger facilities are estimated to be considerably lower. The operating and maintenance cost for a 10 mgd facility similar to the Rosemount facility was estimated to fall in the range of $0.56 to $0.83 per 1000 gal.

III. RECOMMENDATIONS

It is important that the coagulants added in the chemical clarification process be completely dissolved and rapidly mixed with the incoming flow. Subsequent to this, gentle mixing is required to ensure a reasonable degree of flocculation which will, in turn, enhance the removals obtained by sedimentation. Four years of experience at the Rosemount AWTP indicate that mechanical mixers are extremely effective for removing rags and stringy debris from the wastewater flow. It is essential that nonfouling or self cleaning mixing devices be developed so that maintenance effort can be reduced and mixing conditions maintained near optimum.

Although the performance characteristics of the standard automated chemistry modules used to monitor phosphorus concentrations were acceptable, the equipment was costly to operate and maintain. Currently available phosphorus monitors should be evaluated on raw and clarified wastewater because their successful use in control schemes could yield significant saving in terms of coagulant chemicals.

Based on the results at Rosemount and more recent experience at other facilities it appears that there is a need to develop and/or demonstrate a reliable method of protecting mild steel if that material is to be used in contact with activated carbon.

The addition of sodium nitrate can be used to limit the generation of hydrogen sulfide in carbon contactors however the cost of this method of control is significant. Some considerable effort to develop and demonstrate more cost effective means of hydrogen sulfide control is thus justified by the potential savings.

Although the measured organic loading of the carbon columns was relatively constant on a day to day basis the removal efficiency demonstrated significant fluctuations. Work should be continued at the Rosemount facility to determine the cause of these variations in treatment efficiency and thus provide a better understanding of the limitations of activated carbon in treating municipal wastewaters.

Because of the significant costs (capital, operating, and maintenance) involved in carbon transfer and regeneration it would be desirable to develop a method of in situ regeneration to eliminate or at least prolong the interval between thermal regenerations. It was observed by accident at Rosemount that a significant mass of soluble organic material is released from a carbon bed when it is exposed to pH values in the range of 9.5 to 10.5. Leaching with high pH solutions should be investigated.

The data presented in this report and those obtained by other investtigators demonstrate that the working ammonia exchange capacity of the zeolite clinoptilolite is directly related to the regeneration conditions as well as the chemical characteristics of the flow during the service cycle. It is important that the relationships that do exist among the several parameters be better defined so as to allow near optimum design for future treatment facilities.

During the four plus years of operation many engineers have called and visited the Rosemount AWTP in search of information on the performance characteristics and operating problems associated with the unit processes used. Because the operating problems are often related to specific design deficiencies and performance characteristics are affected by many parameters it is not easy to generalize on the applicability of physical chemical treatment or the associated costs. Although several P-C facilities are now operating in the United States there is a dearth of operating information in the technical literature. It is thus recommended that periodic seminars be held to allow design engineers and representatives from all operating P-C facilities to meet and discuss the practical problems encountered by operating personnel. It is anticipated that both groups would benefit from a free and informal exchange of information and possibly lead to more efficient and cost effective treatment.

IV. FACILITY DESCRIPTION

A. GENERAL

The treatment plant is located in the City of Rosemount, Minnesota, approximately 6 miles from the center of the sewered area. The raw sewage flow is transported to the plant through an intercepting sewer which ranges from 27 to 48 inches in diameter and is approximately 5.5 miles long.

The treatment process units are totally enclosed in a 15,000 sq ft steel building, Figure 1. After metering and coarse screening the flow is pumped to one or two parallel process trains each with a design capacity of 300,000 gpd. The flow from the two trains is mixed prior to chlorination and discharged to the Mississippi River (Spring Lake) through a force main approximately one and one half miles long.

The layout of the treatment plant is presented in Figure 2. Each process train consists of a clarifier, two first stage dual media filters, three granular activated carbon columns, two second stage dual media filters, and three zeolite columns. Because of its experimental nature the facility design incorporated flexibility within each process train. One or both sets of filters can be bypassed if desired and the carbon columns can be operated in several modes. In addition the process trains can be coupled in filtered water storage tank No. 2.

Both of the process trains are served by common facilities for sludge storage and hauling, chemical storage, activated carbon regeneration, zeolite regeneration and brine recovery. The major portion of the return flows generated in the plant by backwashing and rinsing are piped to the backwash pit which has a capacity of approximately 15,700 gal and is located between the train 1 and 2 filters below grade. The plant and effluent chlorinators along with two air blowers are located in a separate building approximately 25 feet north of the main building. The boiler, which supplies heat and process steam, and a natural gas powered standby generator are located in a masonry structure directly adjacent to the main building. The maintenance shop is located directly east of the clarifiers.

The laboratory and staff facilities are located in the single story masonry structure south of the main building. Office space for the plant superintendent, maintenance staff and staff associated with the demonstration project is provided in three trailers located near the main building.

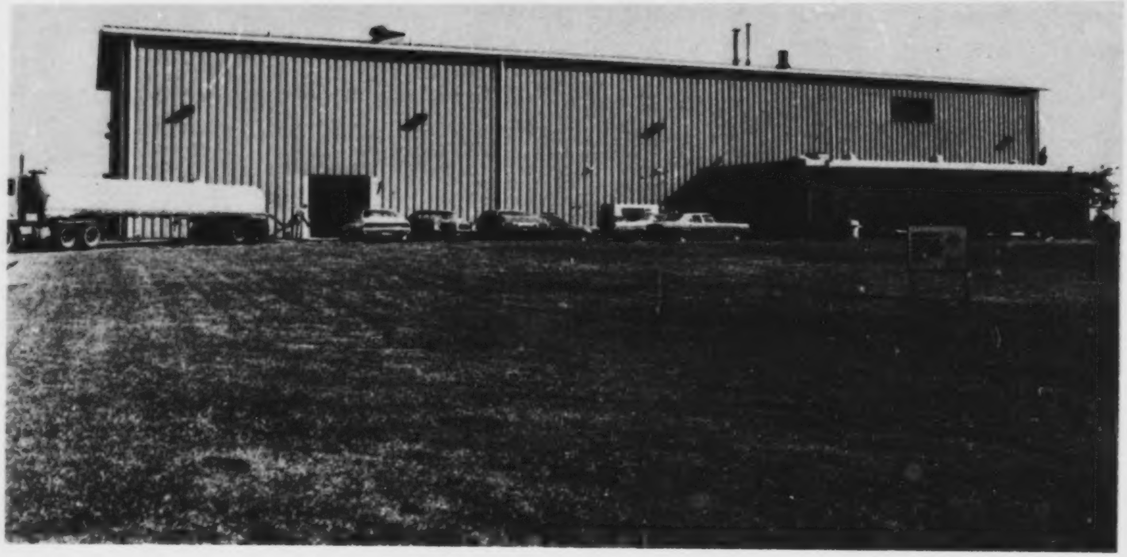


Figure 1. Rosemount Advanced Wastewater Treatment Plant.

1. BAR SCREEN
2. SOLIDS CONTACT CLARIFIERS
3. DUAL MEDIA FILTERS - FIRST STAGE
4. FILTERED WATER STORAGE NO. 1
5. GRANULAR CARBON COLUMNS
6. DUAL MEDIA FILTERS - SECOND STAGE
7. FILTERED WATER STORAGE NO. 2
8. ZEOLITE COLUMNS
9. CHLORINATORS
10. SLUDGE HOLDING TANK
11. CARBON REGENERATION FACILITIES
12. ZEOLITE REGENERATION AND BRINE RECOVERY
13. BLOWERS
14. CHEMICAL STORAGE
15. MAIN ELECTRICAL PANEL
16. BOILER
17. LABORATORY
18. CONTROL ROOM
[illegible] SAMPLING ROOM

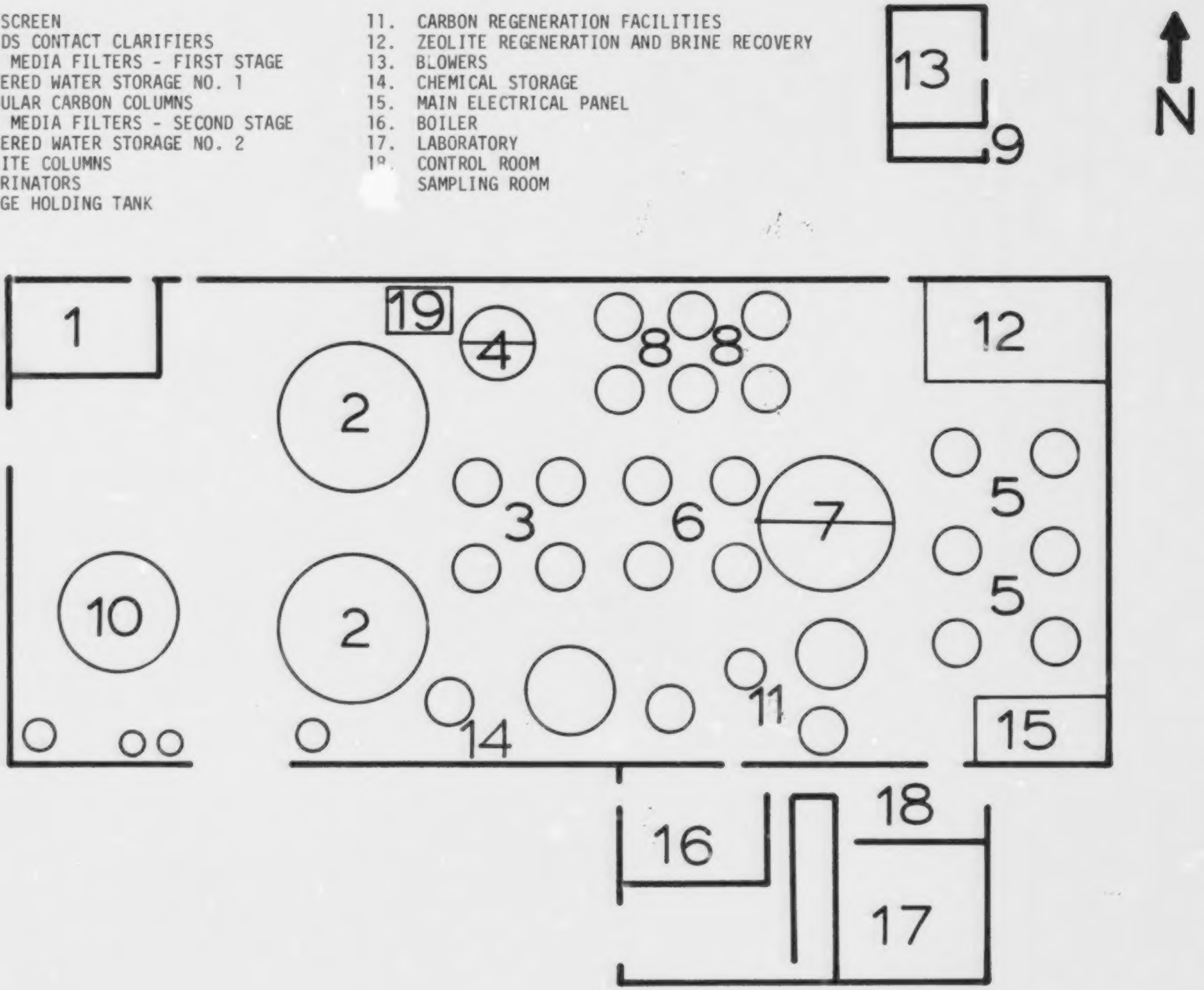


Figure 2. Plant layout.

B. TREATMENT PROCESSES

1. Clarification

A suspended solids contact clarifier is the first treatment unit in each of the process trains. The units were supplied by the Graver Water Conditioning Company and are identified by the trademark Reactivator. The clarifiers are 25 ft in diameter and have a side water depth of 12.67 ft, Figure 3. The clarifiers are constructed of steel and are mounted on concrete pads to elevate the units above grade. The normal water surface elevation is approximately 22 ft above the floor of the plant. The maximum settling area is 424 sq ft. This yields a surface settling rate of 0.5 gpm/sq ft at the design flow of 0.3 mgd per process train.

A variable speed, turbine type impeller is used to mix the treatment chemicals with the sewage flow. The impeller also effects some sludge recycle up through the draft tube. Each unit is equipped with a skimmer and grease box as well as rotating sludge plow. Sludge is removed from the clarifier on a flow proportional basis and discharged to a holding tank.

The clarifier overflow is collected along a peripheral launder and is discharged to a baffle chamber where sulfuric acid is added for pH control as required. The neutralized flow is, in turn, discharged to the filter influent pipes through two triangular weirs.

2. Filtration

Both the first and second stage filters are of the Monoscour design as supplied by Graver Water Conditioning Company. Each filter vessel is 8 ft in diameter, Figure 4. The hydraulic loading rate at design flow is 2 gpm/sq ft. The filter media consist of approximately 24 in. of anthracite coal with an effective size of 1.1 mm and 12 in. of silica sand with an effective size of 0.45 mm. The filter media are supported on the underdrain structure without the aid of any gravel subfill. The underdrain system consists of a steel false bottom fitted with 126 molded plastic strainers. Each of the strainers is covered with a plastic funnel, Figure 5.

A backwash water storage compartment is located above each filter and is integral with the filter vessel. Each of the storage compartments has a capacity of approximately 3000 gal. The second stage filter effluent is used for filter backwash. The backwash compartments of the first stage filters are connected to the compartments of the second stage filters and are filled by gravity flow. With the exception of the flow through the backwash compartment in the second stage filters as illustrated in Figure 4 the first and second stage filters are identical.

During normal operation the filter influent enters through valve A, flows around the baffle plate, through the media and underdrains, through valve W, up through the backwash water storage compartment and over the discharge weir. The backwash procedure is initiated manually although a timer is available for automatic initiation. The backwash sequence is automatically controlled,

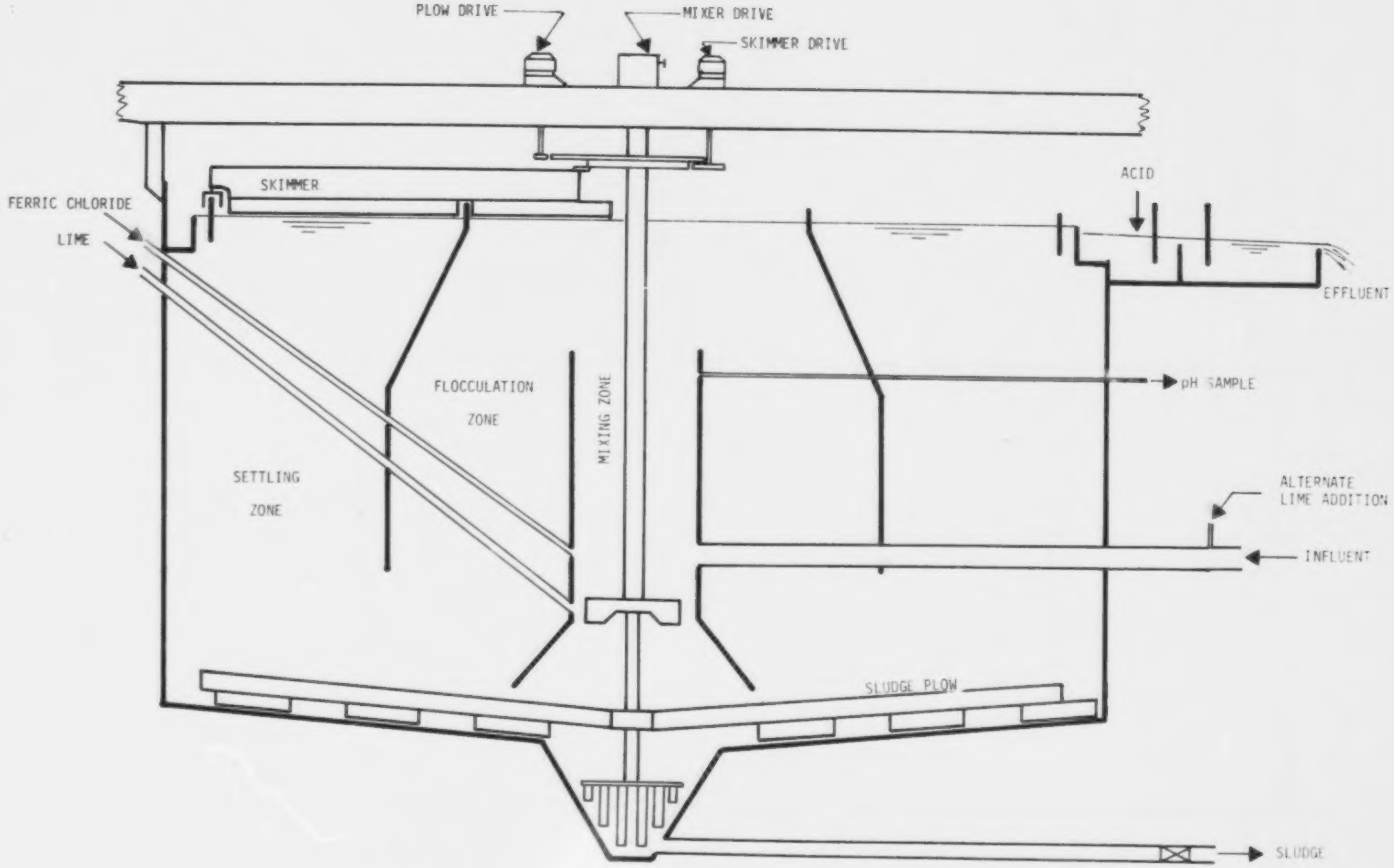

PLOW DRIVE
MIXER DRIVE
SKIMMER DRIVE
SKIMMER
FERRIC CHLORIDE
LIME
ACID
EFFLUENT
FLOCCULATION
ZONE
MIXING ZONE
pH SAMPLE
SETTLING
ZONE
ALTERNATE
LIME ADDITION
INFLUENT
SLUDGE PLOW
SLUDGE

Figure 3. Profile of solids contact clarifier.

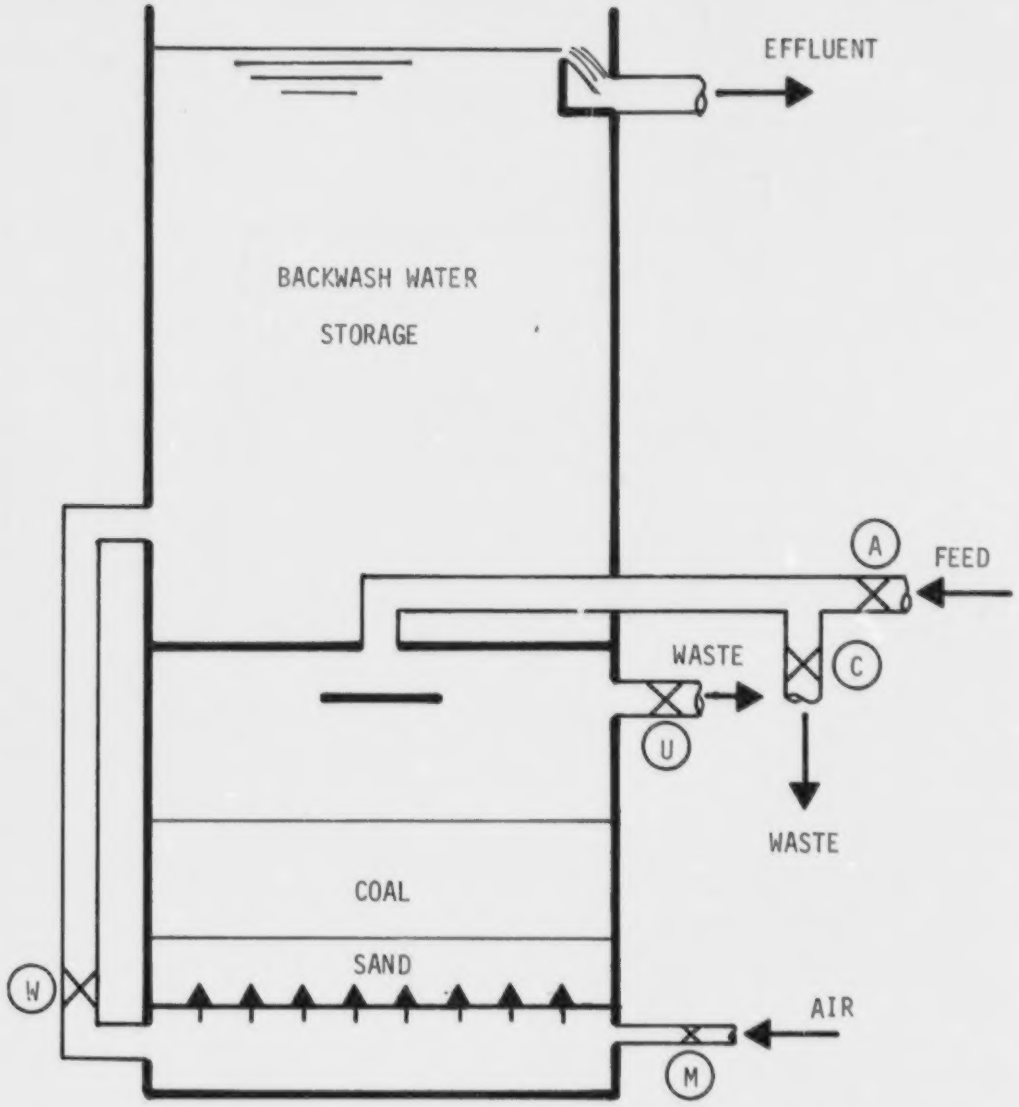

EFFLUENT
BACKWASH WATER
STORAGE
A
FEED
WASTE
C
U
WASTE
COAL
SAND
W
AIR
M

Figure 4. Schematic of second stage dual media filter.

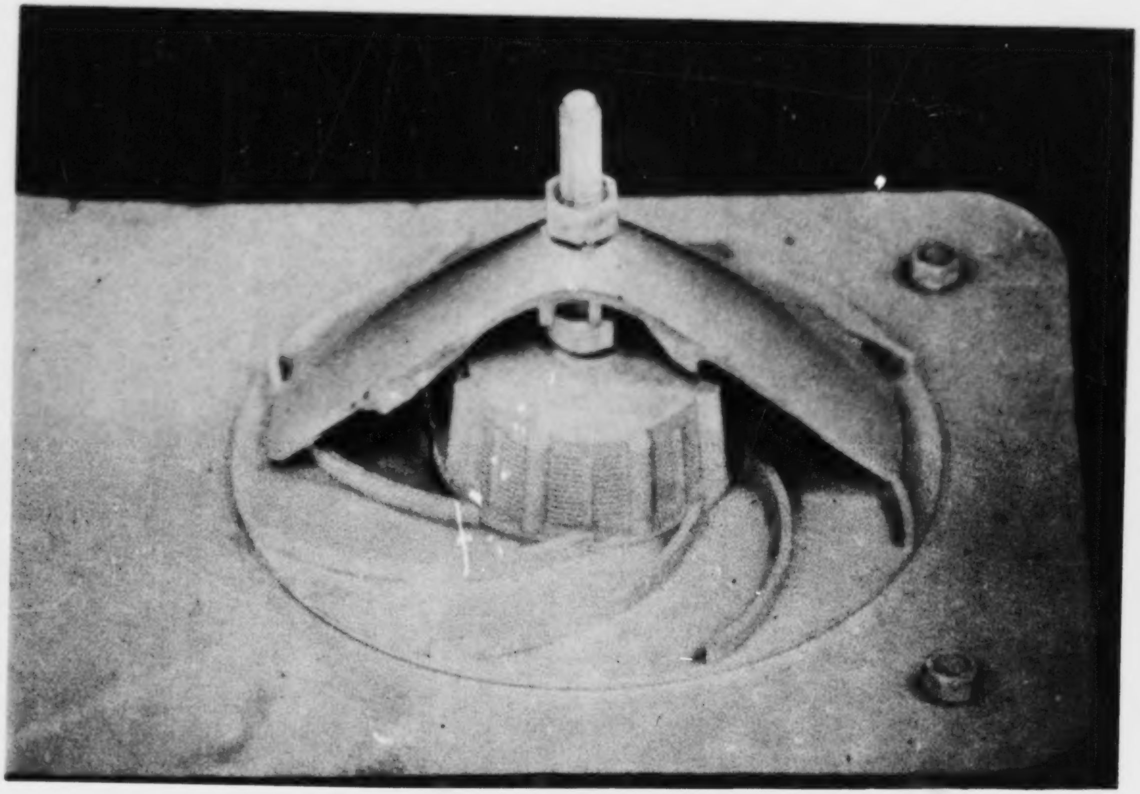


Figure 5. Underdrain strainers.

once initiated, by a multicam timer which energizes the appropriate pneumatic valve operators, TABLE 2.

TABLE 2. FILTER BACKWASH SEQUENCE

Step	Valves Open	Time-min
Draindown	U	5
Air scour	M,C	5
Settle bed	C	6
Backwash	W,C	5
Settle bed	-	3.5
Service	A,W	Variable

The air scour rate is approximately 5 cfm/sq ft and the backwash rate is approximately 16 gpm/sq ft. The filter controls are interlocked to allow only one filter backwash to be in progress per train at any time.

3. Carbon Adsorption

The carbon columns are steel vessels eight ft in diameter and approximately 27 ft high - Figure 6. A nominal depth of 12 ft of granular activated carbon is maintained in each column. The carbon, Westvaco NuChar WV-G, 12 x 40 mesh, is supported on an underdrain system identical to that used in the filters. The system was designed to allow the utilization of several operating modes. Any two of the three columns, per train, may be operated in the following modes: series upflow, series downflow, parallel downflow and primary upflow secondary downflow. In the upflow mode the feed enters the upflow header and is distributed to eight laterals. The laterals are constructed of 3 in. stainless steel pipe with 3/8 in. diameter holes 2 1/2 in. on center along the bottom. At design flow the average carbon contact time, based on empty bed volume, falls between 30 and 40 min depending on the volume of plant recycle flow.

The carbon beds can be backwashed to remove debris and limit the accumulation of biomass. The backwash procedure is initiated manually. Once started however the following sequence is completed automatically: drain to air scour level, stop drain and vent, air scour, settle bed, and backwash. The air scour rate is fixed at approximately 5 cfm/sq ft however the backwash rate and the times for each step in the sequence are controlled by the plant operator.

When the desired organic removal cannot be maintained across the carbon columns, one of the columns is taken out of service and replaced with a column

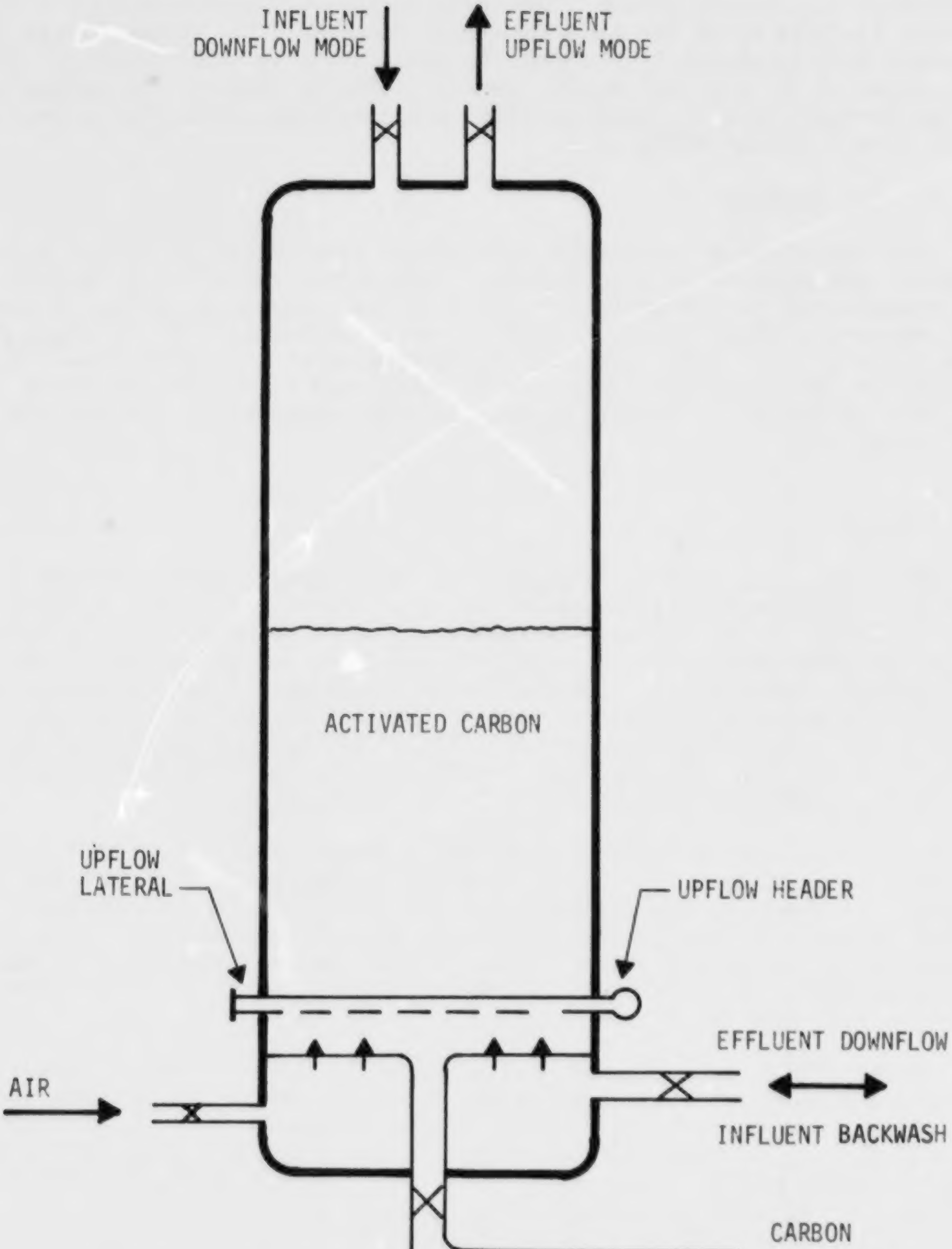

INFLUENT
DOWNFLOW MODE
EFFLUENT
UPFLOW MODE
ACTIVATED CARBON
UPFLOW
LATERAL
UPFLOW HEADER
EFFLUENT DOWNFLOW
INFLUENT BACKWASH
AIR
CARBON
TRANSFER

Figure 6. Schematic of activated carbon column.

containing regenerated carbon. The spent carbon is transported to the regeneration facility by an air aided hydraulic transfer. The column is put in the backwash mode to partially fluidize the bed and air is supplied at the top of the column at 15 to 20 psi as the transfer valve is opened. The carbon is forced through the 3 in stainless steel transfer line and is discharged to the spent carbon holding tank.

4. Ion Exchange

The ion exchange columns are cylindrical steel pressure vessels 8 ft in diameter and approximately 15 ft high. Each column contains a 6 ft depth of the natural zeolite clinoptilolite (20 x 50 mesh) supported on gravel subfill. The underdrain structure consists of a central collection hub and eight pipe laterals with 1 in. holes cut a 45° on both sides of the bottom line. One or two, of the column per train, can be maintained in the service cycle but only in the series downflow mode. Zeolite regeneration is conducted in the upflow mode.

C. PUMPING

The raw sewage enters the plant 17 ft below grade, passes through a Parshall flume, mechanically cleaned bar screen with 1 in. clear spacing and discharges to the wet well approximately 25 ft below grade. All backwash water and other wastewaters generated in the plant are discharged to the wet well via the backwash pit. The raw sewage and recycle flows are pumped to one or both clarifiers. The raw pumping station consists of 3 variable speed pumps each with a discharge capacity of 575 gpm against a total dynamic head of 48 ft. The pump speed and number of pumps in service are automatically controlled based on the water elevation in the wet well.

The carbon column pumps, two per train, are located on the east side of filtered water storage tank No. 1, Figure 2. These pumps draw from the storage tank and discharge to the carbon column inlet manifold. Each pump has a discharge capacity of 250 gpm against a total dynamic head of 55 ft. The carbon column pumps operate a constant speed. A flow control valve, located downstream of the carbon columns, is driven by a level controller in filtered water storage tank No. 1.

The effluent pumps, two per train, draw from filtered water storage tank No. 2 and discharge to the inlet of the zeolite columns. These pumps operate at constant speed and have a capacity of 250 gpm each against a total dynamic head of 85 ft. The rate of flow to the zeolite column is controlled by a float operated recycle valve which maintains a fixed level in filtered water storage tank No. 2.

Water for backwash of the carbon beds and rinse of the zeolite beds is pumped from filtered water storage tank No. 2 with one of the two backwash pumps. Each of the constant speed pumps has a capacity of 750 gpm against a total dynamic head of 51 ft. A flow control loop consisting of a control valve, two mode controller and a flow meter allows the operator to set point and control the backwash rates.

D. FLOW MEASUREMENT

The raw sewage flow is metered with a six in. Parshall flume equipped with a bubbler. The pneumatic signal is used to operate a local readout and is converted to a current signal (4-20ma) for transmission to the plant control room. Here the flow signal is integrated, totalized and recorded.

The process flows are metered at three points in each train; the discharge from the clarifier, pumped discharge to the carbon columns and the pumped effluent flow. After acid addition and mixing in the baffle chamber the clarifier discharge flows through a weir box fitted with two triangular weirs. The head on the weirs is determined with a bubbler. A local readout is provided for the pneumatic signal and the output from a pressure to current transducer (4-20ma) is conducted to the plant control room. The flow signals from both clarifiers are integrated, totalized and recorded. Since all backwash and wastewaters generated in the plant are returned to the plant wet well the recycle rate within the plant is equal to the difference of the raw and clarifier flows for a given time period.

The flow metering system for the discharge of both the carbon column and effluent pumps consists of; flow tube, differential pressure transmitter, integrator and totalizer.

E. CHEMICAL FEED SYSTEMS

1. Lime

Lime slurry is used to elevate the pH in the clarifier as required. The lime delivery system consists of a lime storage silo with a capacity of approximately 30 tons hydrated lime, a slurry mixing tank with a capacity of 1650 gal, slurry transfer pump, a slurry storage tank with 8000 gal capacity and four, variable stroke, diaphragm chemical feed pumps. The lime slurry is made up in the range of 2.5 to 5% on a weight basis. The chemical feed pumps have a capacity of 5 gpm each and are controlled to maintain a flow proportional lime feed rate.

2. Ferric Chloride

The ferric chloride feed system is illustrated in Figure 7. Liquid ferric chloride (approximately 40% $FeCl_3$) is delivered to the plant and stored in a 6500 gal fiberglass tank. The solution is transferred to day tanks, consisting of 55 gal polyethylene drums, on a daily basis. Four, variable stroke, diaphragm chemical feed pumps are used to deliver the ferric chloride to the clarifiers on a flow proportional basis. The pumps have a capacity of 15 gph each. All of the chemical feed pumps used at the plant are similar to those illustrated in Figure 7.

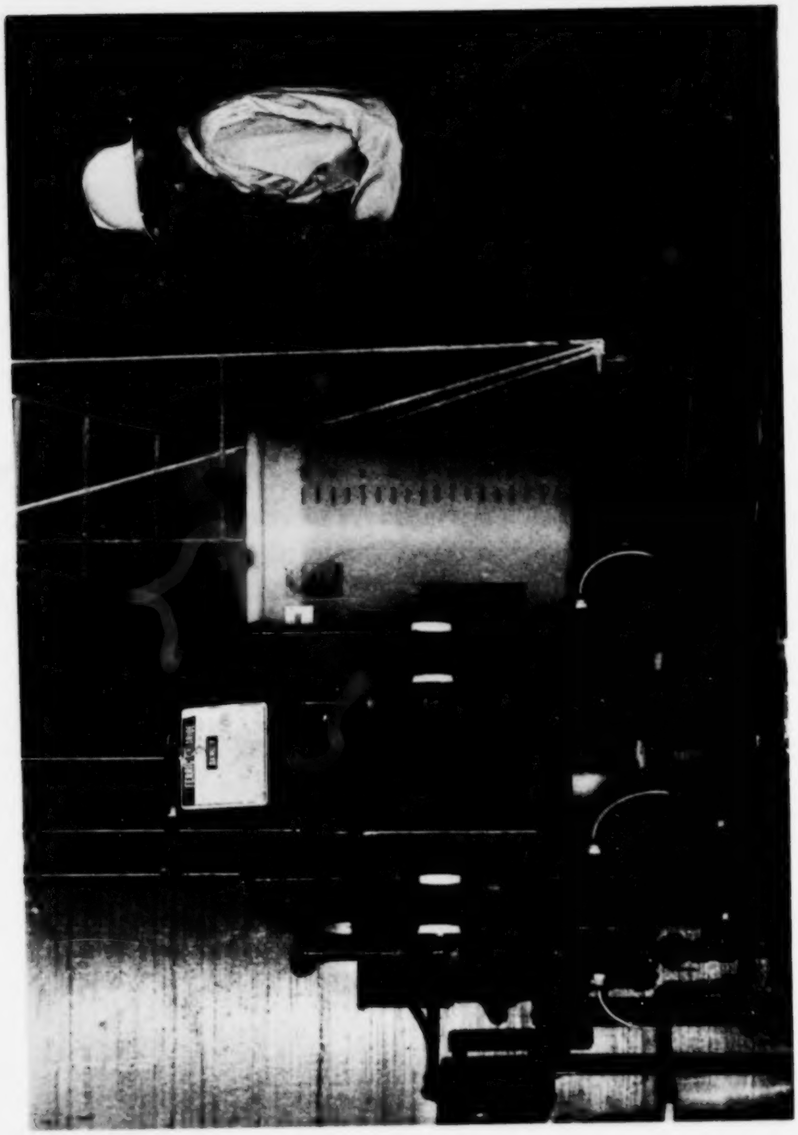


Figure 7. Ferric chloride feed system.

3. Polymer and Nitrate

An anionic polymer was used as a coagulant aid in the clarifier during the early months of the evaluation. The polymer was delivered in liquid form and diluted to a concentration of 0.25% or 0.75% in one of two, 300 gal, fiberglass mixing tanks. Four, variable stroke, diaphragm chemical feed pumps were used to feed the solution to the clarifier on a flow proportional basis. The pumps have a capacity of 10 gph each. Polymer usage was terminatec November 16, 1975.

In early January 1976 this system was modified to feed sodium nitrate solution to the suction side of the carbon column pumps. Sodium nitrate is delivered in 100 lb bags and the solution is made in the fiberglass tanks.

4. Sulfuric Acid

Concentrated sulfuric acid is delivered to the plant site by tank truck and stored in a 4400 gal mild steel tank. The acid is pumped directly from the storage tank to the clarifier as required for pH control. Four, variable stroke, chemical feed pumps are used to deliver the acid based on the output of a pH controller. The capacity of each pump is 10 gph.

5. Chlorine

Chlorine is supplied to the plant in one ton cylinders. Two standard vacuum operated chlorinators with manual rate control are used to deliver chlorine solution to the effluent and at several points in the plant. The plant effluent is served by a unit with a capacity of 100 lb/day. The in-plant chlorine is supplied by a unit with a capacity of 400 lb/day.

6. Caustic

Sodium hydroxide solution, approximately 50% NaOH, is used to adjust the pH of the brine used for zeolite regeneration. The solution is delivered to the plant by tank truck and stored in a heated 800 gal mild steel tank. Caustic is discharged to the brine storage tanks through manually operated valves.

F. ACTIVATED CARBON REGENERATION

The carbon regeneration system is illustrated in Figure 8. Spent carbon is discharged to the dewatering tank where the moisture content is reduced to approximately 50% by gravity drainage. A variable speed screw conveyor delivers the moist carbon to the 54 in diameter refractory lined furnace. The four hearth furnace is equipped with a wet scrubber and afterburner to control stack emissions. Natural gas is used to fuel four burners - two on hearth No. 2 and one each on hearth No. 3 and 4. Steam is injected into the furnace on hearth No. 4 at a controlled rate during thermal regeneration.

The quenched carbon slurry is transported to a storage tank, similar to the on-line carbon columns, with an educator. The regenerated carbon is transferred from the storage tank to one of the on-line columns by the transfer

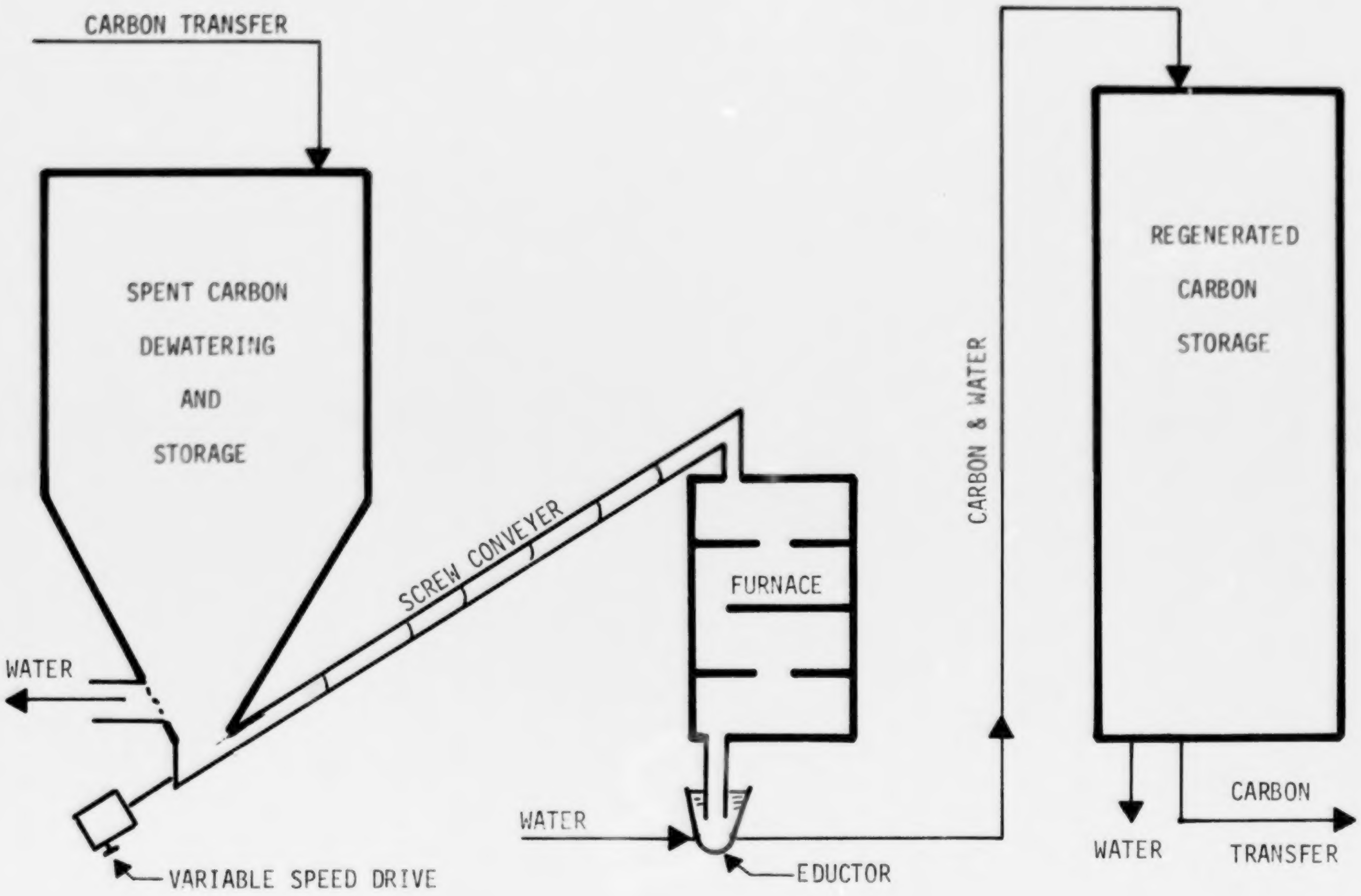

CARBON TRANSFER
SPENT CARBON DEWATERING AND STORAGE
WATER
VARIABLE SPEED DRIVE
SCREW CONVEYER
FURNACE
WATER
EDUCTOR
CARBON & WATER
REGENERATED CARBON STORAGE
WATER
CARBON TRANSFER

Figure 8. Schematic of activated carbon regeneration system.

technique previously described.

G. ZEOLITE REGENERATION AND BRINE RECOVERY

1. General

The zeolite regeneration system consists of facilities for contacting the spent zeolite with a sodium chloride brine at a controlled rate and pH. Brine recovery involves calcium and magnesium removal by soda ash and caustic additions, steam stripping the volatile NH_3 and condensing and collecting the ammonia-water vapor mixture. These two operations combine to form, by far, the most complex system in the plant in terms of operation and physical description. The system consists of the following major items, as illustrated in Figure 9.

Brine storage tanks (2) - below grade concrete construction, fiberglass lined, equipped with 5 hp turbine type agitator and steam heating coils, total volume 15,400 gal, normal working volume of approximately 10,300 gal.

Heat exchangers (2) - plate and frame type, titanium plates, 293 sq ft heating surface, design at 150 psi and 250°F.

Heat exchanger/cooler (1) - plate and frame type, titanium plates, 65 sq ft heating surface, design at 150 psi and 150°F.

Stripping column (1) - fiberglass reinforced plastic construction, 3 ft diameter by 18 ft high, 12 trays design at 14 psi and 250°F.

Condenser (1) - finned tube construction, 860 ft, 1 in. diameter, two 5 hp fixed pitch fans, design at 100 psi and 250°F.

Ammonia receiver (1) - steel tank, 6 ft diameter by 8 ft long.

Brine transfer pump (1) - transfer brine to zeolite columns and stripping column, discharge 100 gpm against 285 ft head, FEP teflon lined ductile iron construction, 25 hp drive.

Stripper product pump (1) - transfer brine from bottom of stripper to reclaimed brine storage, discharge 50 gpm against 165 ft head, FEP teflon lined ductile iron construction, 7.5 hp drive.

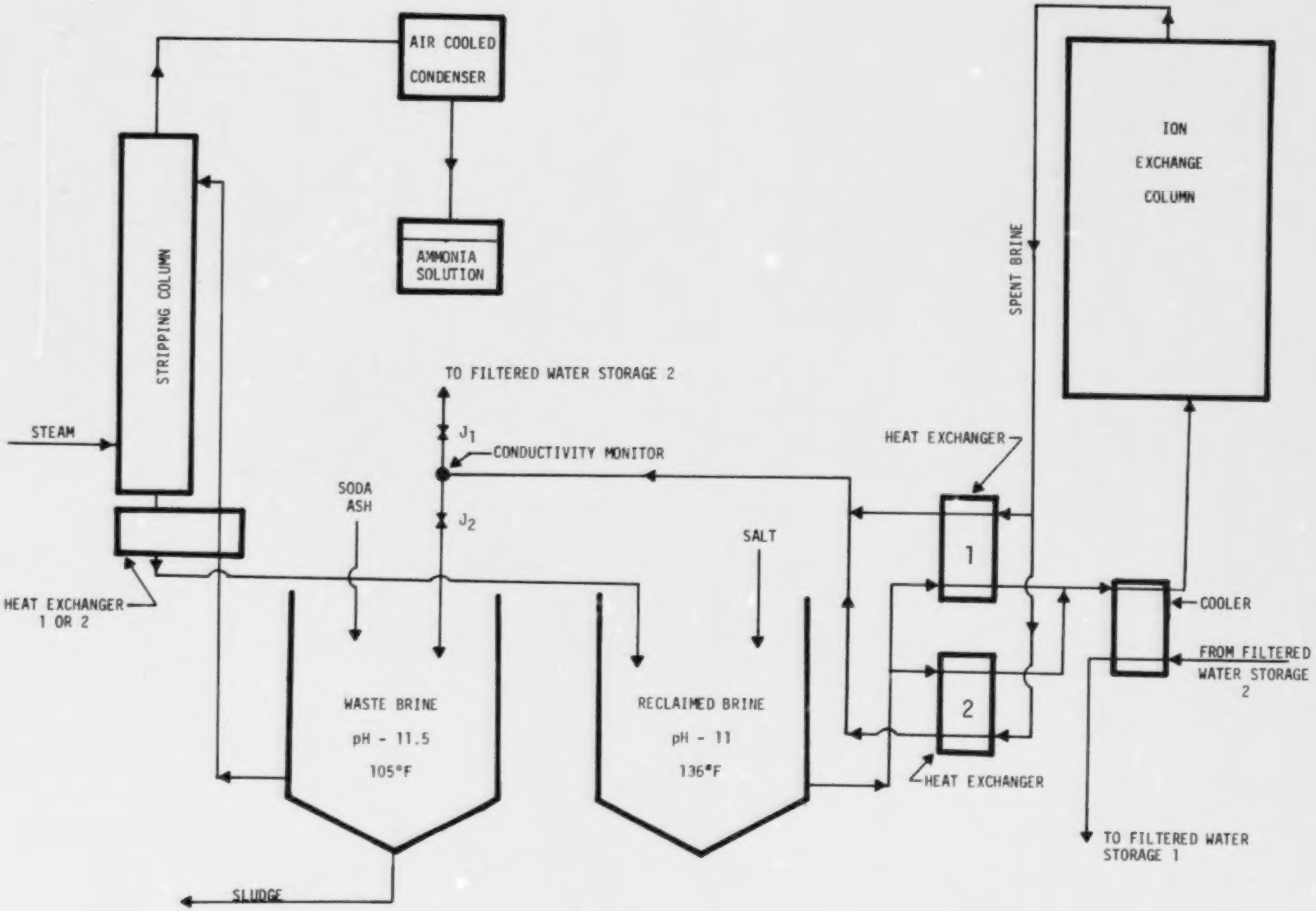

AIR COOLED
CONDENSER
AMMONIA SOLUTION
STRIPPING COLUMN
STEAM
HEAT EXCHANGER 1 OR 2
TO FILTERED WATER STORAGE 2
J_1
CONDUCTIVITY MONITOR
J_2
SODA ASH
WASTE BRINE
pH - 11.5
105°F
SLUDGE
SALT
RECLAIMED BRINE
pH - 11
136°F
HEAT EXCHANGER
1
2
HEAT EXCHANGER
SPENT BRINE
ION
EXCHANGE
COLUMN
COOLER
FROM FILTERED WATER STORAGE 2
TO FILTERED WATER STORAGE 1

Figure 9. Schematic of original zeolite regeneration and brine recovery system.

Caustic storage (1) - steel tank 4 ft diameter by 9 ft high with steam heating coil.

Acid storage (1) - steel tank 2.5 ft diameter by 5 ft high

Fiberglass pipe and rubber lined valves are used throughout this system. The process control panel for the regeneration and recovery system is located in the plant adjacent to the brine storage tanks. The outputs for the require instrumentation - pH, temperature, pressure and conductivity - are located in the panel along with the flow controllers and switches.

The operating procedures for this system changed considerably during the two year demonstration period. The system layout and operating procedures for the initial and final conditions will be presented in the following two sectio All other modifications are described in later sections of this report.

2. Original System

Regeneration is initiated shortly after the zeolite column is taken out of service and consists of the following automated sequence - Figure 9.

(a) Backwash, upflow, at a rate of 8 gpm/sq ft for 10 min.

(b) Settle bed for 5 min.

(c) Pump hot brine through heat exchangers No. 1 and 2, through the cooler (protects zeolite against temperature over 80°F) and up through zeolite bed at a rate of 100 gpm. The process water in the column flows through valve J1 to filtered water storage tank number 2. When the conductivity of this flow exceeds the set point value the conductivity monitor output causes valve J1 to close and J2 to open and the brine is discharged to the waste brine storage tank.

(d) The brine transfer pump shuts down when the brine level in the reclaimed brine tank reaches low level.

(e) Pressurize the zeolite column with air to discharge the brine to the waste brine tank through appropriate valve and pipe arrangement.

(f) Rinse the zeolite bed at a rate of 3 gpm/sq ft, downflow, for 45 min. Discharge the rinse water to filtered water storage tank No. 1.

The spent brine was recovered as follows:

(g) Determine the calcium concentration of the brine.

(h) Adjust the pH to 11.5 with sodium hydroxide solution, add the appropriate weight of sodium carbonate and mix. Allow the $CaCO_3$ and $Mg(OH)_2$ formed to concentrate in the tank bottom and pump to the sludge storage tank.

(i) Pump the desludge brine through heat exchanger 1 or 2 (automatically cycled each regeneration) and discharge to the stripping column at a rate of 50 gpm.

(j) Inject steam into the stripping column at a rate of 1 lb per gal of brine.

(k) Discharge the volatile ammonia and water vapor from the top of the stripper to the air cooled condenser. Collect the condensate in a receiver at a concentration of approximately 1% NH_3-N.

(l) Pump the brine from the bottom of the stripper through the heat exchanger (No. 1 or 2) and discharge to the reclaimed brine tank.

(m) Adjust the pH to 11.0 with caustic or sulfuric acid as required and adjust the sodium concentration to 1.0 N by the addition of NaCl.

3. Final System

The final modification of the regeneration procedure, used during the last 2 months of the study, eliminated the need for heated brine and heat exchangers.

The brine pumping schedule is described in Figure 10 and as follows:

(a) Initial conditions - reclaimed brine tank at level Ⓐ

(b) Pump Ⓨ discharges brine to zeolite column, upflow, at a rate of 100 gpm. The process water in the column exits through valve J1 - Figure 9.

(c) When the reclaimed brine level drops to Ⓓ , valve J1 and valve J2 are manually activated and the spent brine discharges to the waste brine tank.

(d) Pump Ⓨ is automatically shut down when reclaimed brine level reaches level Ⓑ - waste brine now at level Ⓘ.

(e) Pump Ⓩ discharges contents of waste brine tank to reclaimed tank adjusting levels to Ⓙ and Ⓒ respectively - manually controlled.

(f) The pH of the brine in the reclaimed tank is adjusted to 12.0 and solids allowed to settle for 35 min - manually controlled.

(g) Pump Ⓨ discharges brine to zeolite column, upflow, at a rate of 100 gpm.

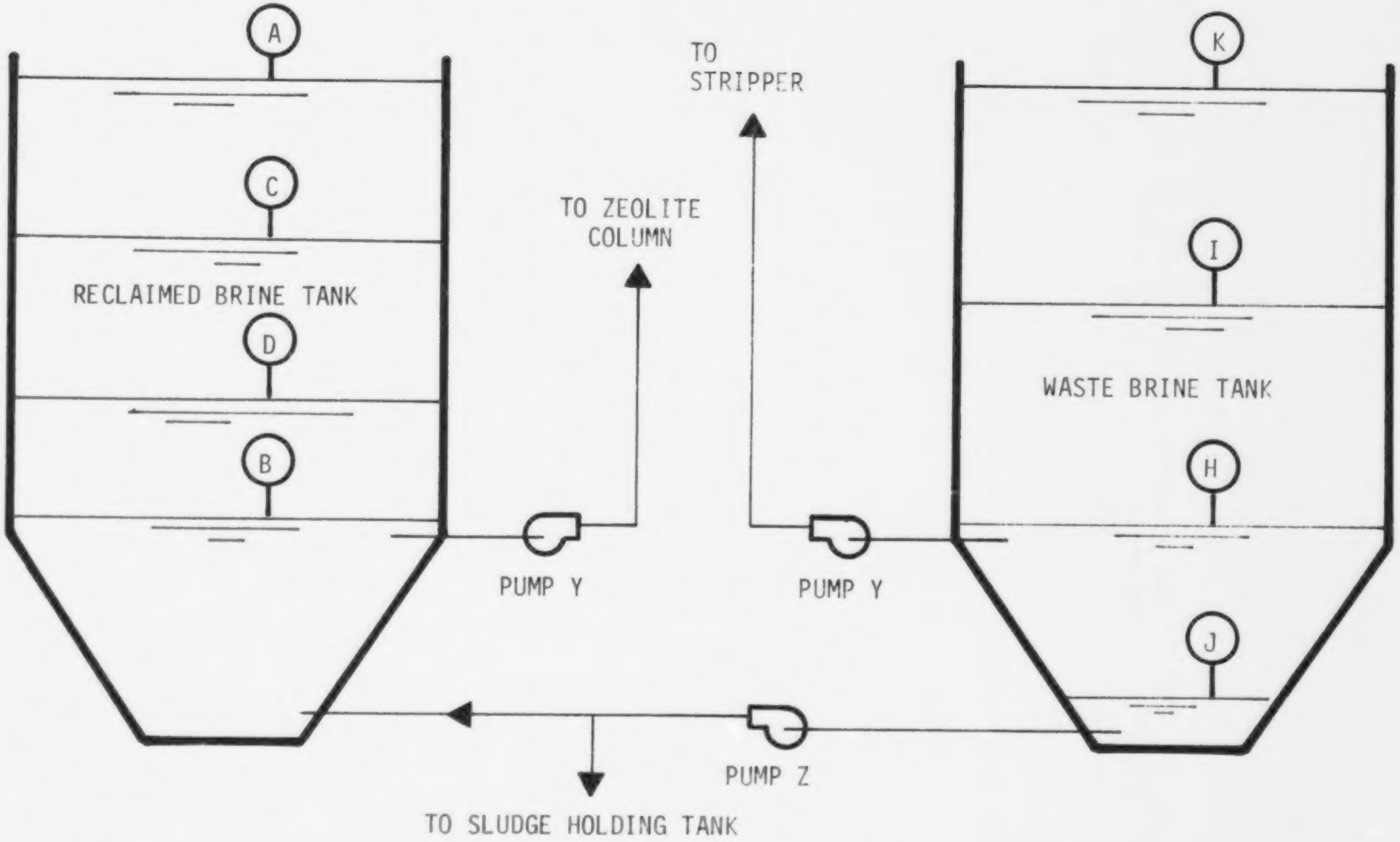

A
C
RECLAIMED BRINE TANK
D
B
TO ZEOLITE COLUMN
TO STRIPPER
PUMP Y
PUMP Y
K
I
WASTE BRINE TANK
H
J
PUMP Z
TO SLUDGE HOLDING TANK

Figure 10. Brine pumping schedule for zeolite regeneration - final conditions

(h) Repeat steps d through g.

(i) Repeat steps d through g.

(j) Repeat step d.

(k) Pressurize the zeolite column with air to discharge brine to waste tank - levels now at Ⓑ and Ⓚ.

(l) Rinse the zeolite bed downflow at 100 gpm for 60 min or until the manually determined pH of rinse water is $\leq$ 8.5. Discharge spent rinse water to headworks via backwash pit.

The spent brine was recovered as follows:

(m) Determine the calcium concentration of the waste brine.

(n) Adjust the pH to 12.0, and the appropriate weight of sodium carbonate added and mixed. Allow the $CaCO_3$ and $Mg(OH)_2$ formed to concentrate in the tank bottom and pump to the sludge storage tank.

(o) Pump waste brine through heat exchanger number 1 and discharge to the stripper at approximately 45 gpm.

(p) Inject steam into the stripper at a rate of 1 lb per gal brine.

(q) Discharge the volatile ammonia and water vapor from the top of the stripper to the air cooled condenser. Collect the condensate in a receiver at a concentration of approximately 1% NH_3-N.

(r) Pump the brine from the bottom of the stripper, through heat exchanger number 1 and discharge to the reclaimed brine tank.

(s) Automatically terminate stripping operation when the waste brine level falls to Ⓗ .

(t) Adjust the pH of the brine to pH 12 and the sodium concentration to 0.5 N by the addition of NaCl.

H. PROCESS CONTROL

1. General

Process control is concentrated at three locations in the plant. The main control panel illustrated in Figure 11 is located in the control room and consists of a graphic panel along with the analog controllers for flow and pH and their associated indicators and recorders. Lights on the graphic panel indicate the status of all treatment process elements - clarifier filters, carbon columns and zeolite columns - along with the status of all pumps and drives and the position of all on-off valves. All of the processes

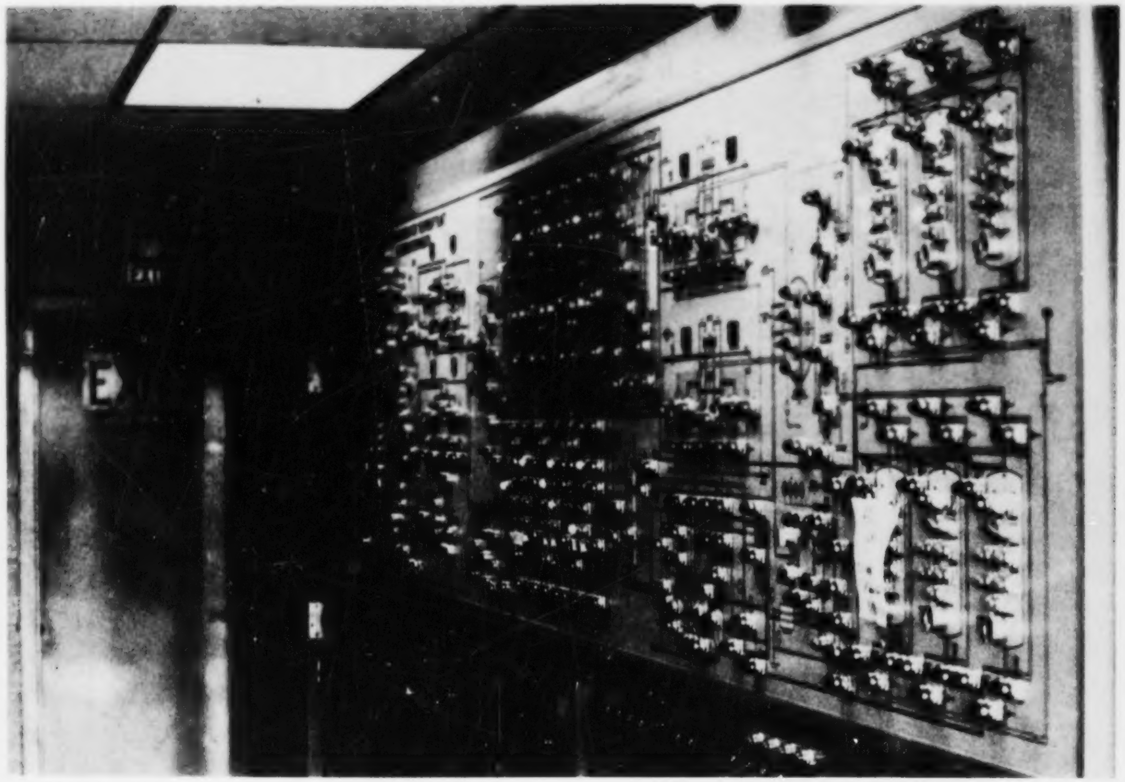


Figure 11. Main process control panel.

with the exception of carbon regeneration and brine recovery are controlled from this main panel. The carbon regeneration and brine recovery operations are controlled at smaller local panels.

2. Clarifier

The flow rate through the clarifier is determined with two triangular weirs located just downstream of the point of acid addition, Figure 3. The flow signal, 4-20 ma, is used to control the addition of lime and ferric chloride and the sludge blowdown rate. The stroke adjusting mechanisms of the chemical feed pumps are operated directly from the flow signal with 4 to 20 ma yielding 0% to 100% maximum stroke respectively. The maximum stroke or capacity of each pump is field adjustable to allow operation over a wide range of concentrations.

The flow integrator outputs a 24 VDC pulse for every 10 gal discharged. This pulse activates the flow totalizer and an automatic reset count controller. When the operator entered count value is reached the counter resets and activates the sludge blowdown cycle. The sludge blowdown is preceded by a water backflush to clear the blowdown line. Both the rinse and the blowdown times are operator entered.

A sample from the draft tube is transported by gravity to a pH monitor located outside the clarifier. The signal generated is transmitted to the pH receiver and recorder on the main control panel. The plant operator records the pH value every hour and adjusts the lime dosage as required. Although the alkalinity of the raw wastewater is relatively constant the varying rates of recycle affect the alkalinity of the clarifier influent and, in turn, the pH at constant lime dose. Precise control is difficult because of the long response time. For this reason the lime dose is not changed unless the pH deviates by more than 0.5 from the set point value.

A sample of the acid neutralized clarifier effluent is transported by gravity to a second pH monitor. The signal is transmitted to the pH receiver and, in turn, to a recorder and PID controller. The controller outputs to the stroke adjusting mechanism of the sulfuric acid feed pump based on the deviation from the pH set point value.

3. Filters

Each filter is equipped with a differential pressure transmitter to monitor the head loss across the filter media. The head loss is indicated on the main control panel. The filters are generally backwashed once each day and backwash is initiated on a time schedule rather than on a head loss basis. This procedure is followed to ensure that most of the recycle flow is returned to the headworks during periods of low raw wastewater flow and thus equalize somewhat the hydraulic loading on the plant. The head loss indicators are used to evaluate the efficiency of the backwash procedure and to determine when special cleansing procedures are required.

4. Carbon Regeneration

The carbon regeneration furnace controls are mounted in a panel adjacent to the furnace. The panel contains the following devices:

(a) on/off/fail switch/indicators for all drives

(b) small alarm status panel

(c) indicating temperature controllers for hearths 2, 3 and 4 and the afterburner

(d) furnace damper controllers

(e) furnace pressure indicator

(f) scrubber differential pressure indicator

(g) rabble arm speed control

Although the furnace was the smallest size available at the time of plant construction its capacity is greater than required. For this reason, along with the fact that the plant is being operated at approximately 40% of design capacity, the furnace is operated for 5-10 day periods only two or three times per year. The furnace is brought up to operating temperature slowly during a 24 to 30 hour period to minimize thermal stresses in the refractory material. This is accomplished by manually adjusting the set points on the temperature controllers.

Carbon starts to discharge from hearth number 4 approximately one hour after initiation of carbon feed. Throughout the regeneration the hearth temperatures are maintained in the following ranges:

Hearth	°F
1	900 to 1050
2	1400 to 1500
3	1550 to 1650
4	1600 to 1700

Steam is added on hearth number 4. The steam addition rate is maintained and regulated by a combination of a pressure regulating valve, rotometer and control valve.

The rates of carbon feed and steam addition are determined by the operator based on the apparent density of the regenerated carbon. The procedure and a description of the apparatus required to determine apparently density are found elsewhere.(2). The apparent density of the virgin carbon loaded in the columns were 0.48 gm/cc. The plant operator samples the regenerated carbon, prior to quench, each hour and determines the apparent density. If the value falls outside the range 0.46 to 0.50 the operator adjusts the carbon feed rate and/or the steam feed rate accordingly. The apparent density is decreased by reducing the carbon feed rate or increasing the steam rate. Composite samples of both the regenerated and spent carbon are constructed from the hourly grab samples. The composite samples are analyzed at a later date to determine the efficiency of the regeneration.

5. Zeolite Regeneration and Brine Recovery

The zeolite beds are put into service at midnight and taken out of service 24 or 48 hours later depending on the mode of operation. Because of time constraints and the limited storage capacity for spent backwash and rinse water, the length of the service cycle is not varied to reflect ammonia loading and breakthrough but rather maintained constant. The regeneration is controlled from the main panel. The regeneration cycle is essentially completely automated however the operator must set the sequence times and flow controllers prior to initiation.

The brine recovery process is controlled at a local panel adjacent to the brine storage tanks. The following instruments are available for control.

(a) brine pH monitors

(b) conductivity monitor

(c) temperature readouts for brine and stripper

(d) flow controllers for brine transfer and steam

(e) stripper liquid level control

The brine recovery process is manually controlled with the aid of two automated flow control loops (brine and steam feed rates) and automatic control of the product level in the stripper bottom. The set point values for the controlled variables are maintained constant from day to day and changed only when problems are encountered such as high head loss in the heat exchanger or high temperature in the ammonia receiver.

V. ANALYTICAL PROGRAM

A. SAMPLING AND ANALYSIS SCHEDULE

1. Treatment Processes

The original demonstration project agreement specified the location and frequency of sampling along with an analysis schedule. The sampling frequency and analysis schedule was modified when the Minnesota Pollution Control Agency issued the NPDES discharge permit for the plant in June 1975. During the project it became obvious that additional sampling and analyses were required to better evaluate the unit processes. The sample identification scheme used at the plant and throughout this report is presented in TABLE 3.

TABLE 3. SAMPLE IDENTIFICATION SCHEME

Sample	Label
Plant influent	Raw
Clarifier influent	Clarin
Clarifier effluent	Clar
First stage filter effluent	DMF_1
Primary carbon column effluent	GCC_1
Carbon column system effluent	GCC
Second stage filter effluent	DMF_2
Influent to primary zeolite column	INAEC
Primary zeolite column effluent	AEC_1
Zeolite column system effluent	AEC
Plant effluent	Eff

All of the above samples were collected on a flow proportional basis. The flow signals used to activate the samplers and the location of the sample taps are summarized in TABLE 4. The raw sewage sample was transported from the wet well area to the sampler on the main floor through iron pipe using a Moyno model WA400 grinder pump. The DMF_1 and DMF_2 samples were delivered to the sampling apparatus through Moyno model FM221 pumps. All other samples described in TABLES 3 and 4 were delivered to the sampling apparatus through PVC pipe and polyethylene tubing by the head available at the sample tap.

The Raw and Clarin samples were collected in a Sanford Products Corp. sampler model TC-2. The samples were collected in one gal polypropylene bottles under refrigation. All other flow composite samples were transported to the 8 ft x 11 ft sampling and monitoring enclosure, Figure 2, and collected in the custom built apparatus described in Figure 12. Each sample flowed through a constant head box constructed of acrylic sheet and discharged through polyethylene tubing to the sample box also constructed of acrylic sheet. A double acting pneumatic cylinder was used to locate the sample discharge above the waste and sample compartments. The sample compartment discharged to a refrigerated 0.5 gal polyethylene bottle through Tygon tubing. The pneumatic cylinder was actuated by a four-way pneumatic solenoid valve. The sample volume collected during a day was controlled by the reset count controller and the adjustable time delay relay. The relays were set to provide sample times of 1 to 2 seconds. The count controllers were set to actuate the sampler once every 1750 to 4000 gal. These combinations yielded sample volumes of slightly less than one gal per day. Clean sample bottles were placed in the refrigerator each day at approximately 0000 hr and 1350 hr. The two sample bottles were stored in the laboratory refrigerator prior to mixing. The routine analysis schedule utilized at the end of the project period is presented in TABLE 5. For the most common sampling frequency, 3 per week, samples were collected on Tuesday, Wednesday and Thursday and analyzed on Wednesday, Thursday and Friday. This schedule was adopted to eliminate laboratory hours on Saturday and Sunday, i.e. 5 day BOD's.

Grab samples of the clarifier blowdown were collected twice each weekday during periods of high and low flow. The total solids concentrations of the two samples are averaged to estimate the solids concentration for each day.

2. Activated Carbon Regeneration

The shift operator collected a sample of the regenerated carbon, prior to quench, every hour the furnace was discharging carbon. After the sample was cooled in a closed metal container a 50 ml volume was placed in the composite bucket. After each furnace run the composite sample, approximately 4000 ml volume, was reduced to approximately 250 ml in volume with a riffle type sample splitter (Humbolt model 3985). Samples of the moist spent carbon were collected every two hours and placed in individual plastic bags. The samples were dried overnight at 103°C and a composite constructed using equal volumes from each grab sample. The following analyses were conducted on the composite samples: apparent density, iodine number, ash and volatiles.

TABLE 4 SAMPLE LOCATION AND FLOW SIGNAL

Sample	Flow Signal	Location
Raw	Raw	Discharge of Parshall flume upstream of bar screen
Clarin	Clarifier	Clarifier inlet pipe outside clarifier wall
Clar	Clarifier	Weir box upstream of effluent weir
DMF_1	Clarifier	Inlet pipe to filtered water storage tank number 1
GCC_1	Carbon Column	Column effluent pipe
GCC	Carbon Column	Upstream of inlet and bypass valves for second stage filters
DMF_2	Carbon Column	Backwash water storage compartments - both filters
INAEC	Effluent	Discharge manifold of effluent pumps
AEC_1	Effluent	Column effluent pipe
AEC	Effluent	Column effluent manifold upstream of Cl_2 addition
Eff	Effluent	Discharge of chlorinator feed pump

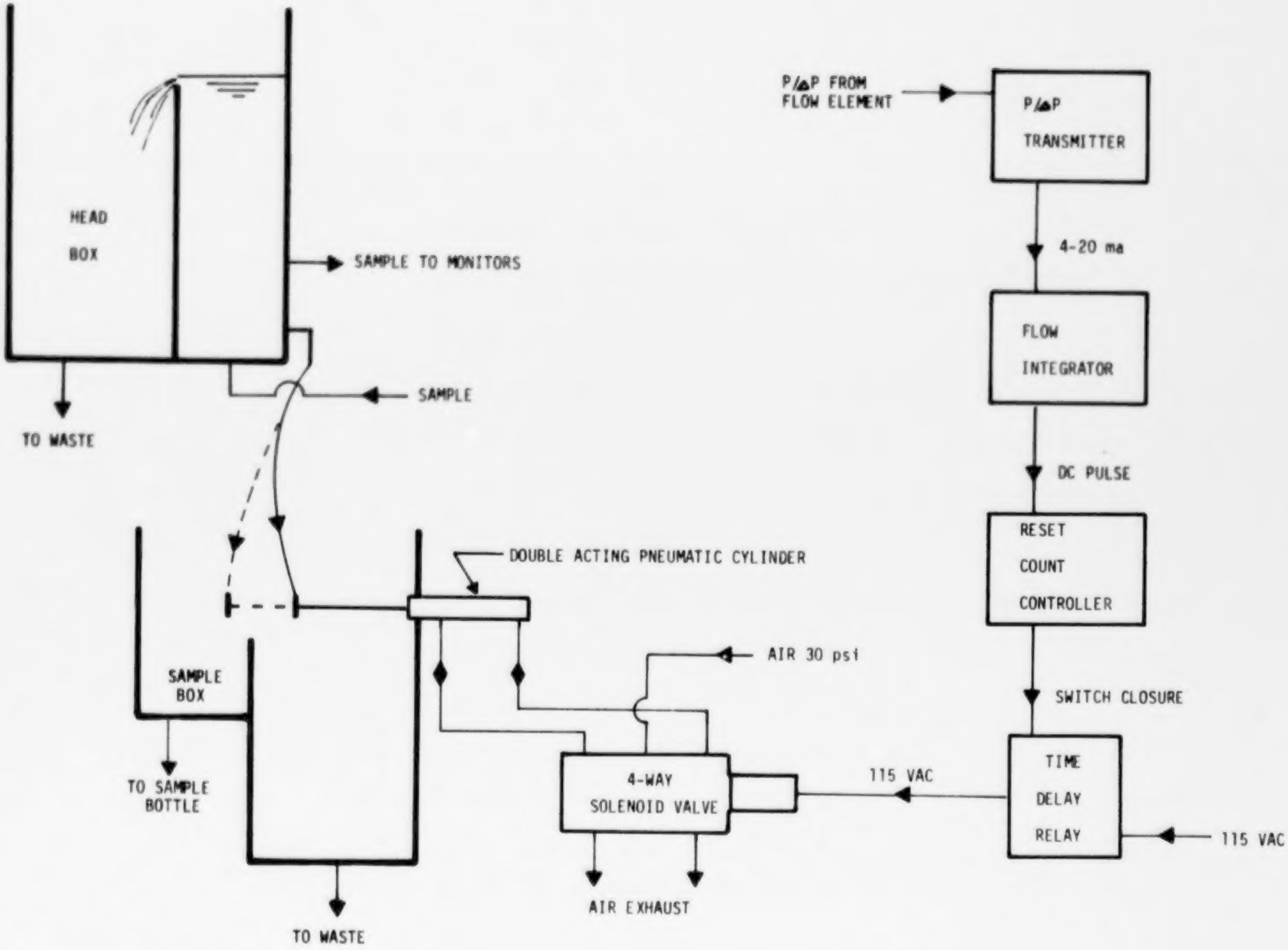

HEAD
BOX
TO WASTE
SAMPLE TO MONITORS
SAMPLE
DOUBLE ACTING PNEUMATIC CYLINDER
SAMPLE
BOX
TO SAMPLE
BOTTLE
TO WASTE
AIR 30 psi
4-WAY
SOLENOID VALVE
AIR EXHAUST
115 VAC
P/ΔP FROM
FLOW ELEMENT
P/ΔP
TRANSMITTER
4-20 ma
FLOW
INTEGRATOR
DC PULSE
RESET
COUNT
CONTROLLER
SWITCH CLOSURE
TIME
DELAY
RELAY
115 VAC

Figure 12. Typical compartment and control of sample collection system.

TABLE 5. ROUTINE SAMPLING AND ANALYSIS SCHEDULE FOR TREATMENT PROCESSES

Parameters	Analyses per Week for Samples Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	EFF
BOD - total	3	3	3	3	3	3	3	-	-	-	3
BOD - soluble	3	3	3	-	3	3	-	-	-	-	3
COD - total	3	3	3	3	3	3	3	-	-	-	3
COD - soluble	3	3	3	3	3	3	-	-	-	-	3
TOC - total	3	3	3	3	3	3	3	-	-	-	3
TOC - soluble	3	3	3	3	3	3	-	-	-	-	3
Suspended Solids	3	3	3	3	3	3	3	-	-	-	3
Turbidity	-	-	3	3	3	3	3	-	-	-	3
Settleable Solids	7	-	-	-	-	-	-	-	-	-	7
Temperature *	7	-	**	**	-	**	**	-	-	**	7**
pH	7	3	3**	3**	3	3**	3**	3	3	3**	7**
Alkalinity	3	3	3	-	-	-	-	-	-	3	3
Chloride	3	3	3	-	-	-	-	-	-	-	3**
Total Solids	1	-	-	-	-	-	-	-	-	-	1
Total Volatile Solids	1	-	-	-	-	-	-	-	-	-	1
Total P	3	3	3	-	-	-	-	-	-	-	3
Total Soluble P	3	3	3	-	-	-	-	-	-	-	3
Total Ortho P	3	3	3	-	-	-	-	-	-	-	3
Soluble Ortho P	3	3	3	-	-	-	-	-	-	-	3
Ammonia-N	3#	-	-	-	-	-	-	#	#	#	#
Kjeldahl-N	3	3	3	-	-	-	-	3	-	3	3
Hardness	1	-	-	-	-	-	-	-	-	-	1
Calcium	1	-	-	-	-	-	-	1#	1#	1#	1
Potassium	-	-	-	-	-	-	-	#	#	#	-
Sulfide - H_2S	-	-	-	##	##	##	-	-	-	-	-
Dissolved Oxygen *	-	-	-	-	-	-	-	-	-	-	7
Chlorine Residual *	-	-	-	-	-	-	-	-	-	-	7
Fecal Coliform *	-	-	-	-	-	-	-	-	-	-	5
Heavy Metals	+	+	+	+	+	+	+	-	-	+	-

* - grab sample, ** continuously monitored, # analyzed every day zeolite columns in service, ## variable frequency

+ - monthly

3. Zeolite Regeneration and Brine Recovery

A number of grab samples were collected during each regeneration and recovery sequence. Sample identifications and locations are summarized in TABLE 6. The analytical schedule is presented in TABLE 7. The shift operator collected the samples and conducted the analyses required for process control. The ammonia and potassium analyses were conducted by the laboratory staff.

TABLE 6. SAMPLE LOCATION - ZEOLITE REGENERATION AND BRINE RECOVERY

Sample	Location
Reclaim 1	Reclaimed brine tank prior to regeneration
Waste 1	Waste brine tank after brine passed through zeolite bed 1 time
Reclaim 2,3,4	Reclaimed tank prior to 2nd, 3rd and 4th pass
Waste 2,3,4	Waste brine tank after brine passed through zeolite bed 2nd, 3rd and 4th times
Desludged	Discharge of brine transfer pump (enroute to stripper)
Stripped	Reclaimed brine tank prior to adjusting pH and Na

TABLE 7. ANALYTICAL SCHEDULE - ZEOLITE REGENERATION AND BRINE RECOVERY

Sample	pH	NH_3-N	Na	Ca	K
Reclaim 1	x	-	x	-	x
Reclaim 2,3,4	x	-	-	-	-
Waste 1,2,3	x	-	-	-	-
Waste 4	x	x	x	x	x
Desludged	-	-	-	x	-
Stripped	x	x	x	-	-

B. PROCEDURES

The total phosphorus, Kjeldahl nitrogen and heavy metals analyses were conducted at the MWCC central laboratory. The samples for P and N analyses were acidified, refrigerated and transported to the laboratory once each week. The samples for metals analyses were acidified and transported to the central lab where they were refrigerated prior to analysis. All other analyses were conducted in the plant laboratory by the laboratory staff except as previously noted.

Laboratory methods and equipment are summarized in TABLE 8. The soluble fractions for the BOD, COD and TOC analyses are defined as that portion passing through a Reeve Angel 934 AH glass fiber filter. Soluble phosphorus is defined as phosphorus passing a 0.45u, distilled water washed, membrane filter.

All flow composite samples were collected on a calendar day basis. The analyses conducted at the Rosemount laboratory were generally completed the day following sample collection.

C. CONTINUOUS MONITORING

Prior to the initiation of the two year study it was anticipated that a limited amount of on line monitoring would be required to better define process performance characteristics. The description of the monitoring equipment used is summarized in TABLE 9. All of the monitoring equipment was located in the sample room - Figure 2. The temperature, pH, conductivity and chloride ion measurements were made on a continuous basis. The sample streams monitored are presented in TABLE 10.

Two sample manifolds were constructed to allow semi-continuous monitoring, for TOC & turbidity and phosphorus & ammonia, of up to 6 sample streams each. Sample was delivered from the head box, Figure 12, to a polyethylene header through air operated pinch tube valves. These valves were closed and opened by applying and relieving pressure (air @ 30 psi) to the rubber membranes. The air flow was controlled by 3-way solenoid valves which were, in turn, controlled by a step programmer (ATC model 1800) and timer (ATC model 333). Samples were cycled to the TOC and turbidity monitors at approximately 20 min intervals and to the phosphorus and ammonia monitor at 10 min intervals. The numbers and identities of the samples cycled to the monitors varied considerably and were dependent on the mode of operation and plant or process operating problems encountered during a given time period.

TABLE 8. SUMMARY OF ANALYTICAL PROCEDURES

Analysis	Method and Equipment	Reference - Page
BOD	Direct dilution, Weston & Stack 330 DO meter	3 - 489
COD	Manual reflux	3 - 495
TOC	Astro Ecology Model 1200	4 - 236
Suspended solids	Reeve Angel 934AH glass fiber filter	3 - 537
Turbidity	Hach model 2100 A	4 - 295
Settleable solids	Imhoff cone	3 - 539
Temperature	Laboratory grade thermometer	3 - 348
pH	Fisher Accumet 230 pH meter	3 - 500
Alkalinity	Total, methyl orange	3 - 370
Chloride	Mercuric nitrate	3 - 97
Total solids	103°C oven	3 - 535
Total volatile solids	550°C furnace	3 - 536
Total phosphorus	Automated perchloric acid digestion - Technician Auto Analyzer	- -
Total ortho P	Stannous chloride, Bausch & Lomb Spectronic 70	3 - 530
Ammonia N	Direct Nesslerization, Bausch & Lomb Spectronic 70	3 - 226
Kjeldahl N	Automated perchloric acid digestion - Technician Auto Analyzer	- -

TABLE 8 Continued

Analysis	Method and Equipment	Reference - Page
Hardness/Calcium	EDTA titration,	3 - 179, 84
Potassium	Orion electrode model 93-19, Orion specific ion meter 407A	5 -
Sodium	Fisher electrode model 13-639-20, Orion specific ion meter 407A	- -
Sulfide	Hach sulfide test kit model HS-6	- -
Dissolved Oxygen	Weston & Stack 330 DO meter	4 - 56
Chlorine Residual	Wallace & Tiernan amperometric titrator	3 - 112
Fecal coliform	0.45 u membrane filter, 44.5 C water bath	3 - 694
Heavy Metals	Atomic absorption - Perkin Elmer 103/305	3 -
Apparent Density		2 - B23
Iodine Number	5 or 6 point isotherm, Spex Industries, Inc. Ball Mill model 8000-11, Burrell Shaker model 75	6 - 249
Carbon Ash	750° C furnace, Vicor crucible	7 - 373
Carbon Volatiles	950° C furnace, nickel crucible	8 - 375

TABLE 9. MONITORING EQUIPMENT

Parameter	Hardware
Tmperature	Type T thermocouple, Leeds and Northup (L & N) Speedomax W multipoint recorder
pH	L & N 7779 flow through transmitter mounting with electrodes, L & N 7073 multi-channel pH receiver, L & N Speedomax W multipoint recorder
Specific conductance	Uni-Loc flow through cell No. 400176, Uni-Loc conductivity transmitter No. 711, Soltec recorder No. B261A
Chloride ion	Uni-Loc flow through sensor No. 320, Orion chloride electrode No. 94-17, Uni-Loc selective ion transmitter No. 1032, Soltec recorder No. B261A
TOC	Astro Ecology No. 1000 total organic carbon analyzer
Turbidity	Hach No. 2411 Surface Scatter turbidimeter Soltec recorder No. B161
Phosphorus & Ammonia	Technician Autoanalyzer II - porportioning pump II, 2 single channel colorimeters, dual pen recorder, 2 continuous type filters, ammonia cartridge, total inorganic phosphate cartridge, high temperature bath

TABLE 10. CONTINUOUS MONITORING SCHEDULE

	Sample Stream					
	Clar	DMF_1	GCC	DMF_2	AEC	EFF
Temperature	x	x	x	x	x	x
pH	*	x	x	x	x	x
Specific conductance	-	-	-	-	-	x
Chloride ion	-	-	-	-	-	x

* recorded at main process control panel

VI. STAFF

The permanent staff involved in operating, maintaining and evaluating the Rosemount AWTP during the two year study consisted of 17 persons as presented in TABLE 11. Two additional operating attendants were used during the summer months.

TABLE 11. PERMANENT STAFF POSITIONS

Position	number	shift
Operator	4	all
Assistant operator	4	all
Operating attendant	2	day
Electrician	1	day
Pipe fitter	1	day
Secretary	1	day
Chemist	1	day
Laboratory technician	1	day
Superintendent	1	day
Project manager	1	day

The staff during the afternoon (3 pm - 11 pm) and midnight (11 pm - 7 pm) shifts consisted of an operator and an assistant operator. The day shift staff during the normal work week consisted of an operator, assistant operator, two operating attendants and the other permanent staff. On weekends the plant was staffed with two-man shifts only. Essentially all of the maintenance work was conducted by the plant staff.

During the winter months the sludge generated at the plant was hauled to another MWCC facility for disposal by a private contractor. During the summer months of 1975 and 1976 the sludge was injected on land adjacent to the plant. All sludge generated during 1977 was hauled to other disposal facilities. The injection work during the 1975 season was conducted by MWCC personnel not listed above. During the 1976 season the plant staff carried out approximately one half of the injection tasks.

Janitorial services for the control room, locker room, laboratory and offices were provided by a private contractor.

VII. COST ACCOUNTING

A separate accounting system was established at the Rosemount facility in June, 1974 to develop cost data for each of the processes utilized. The plant was divided into 17 cost centers as illustrated in TABLE 12. Physical boundaries define 15 of the cost centers. One cost center, Laboratory Services, is defined by function and the remaining cost center, Indirect Services, is used to identify charges that cannot reasonably be identified with direct plant operation.

During normal operation, the three major costs for a plant of this type are labor, chemicals, and power. All MWCC employees at the plant completed two time cards on a biweekly basis, the regular time card for pay purposes and a supplemental time report to identify hours in the various cost centers. Time estimates were made in half hour increments.

Chemical usage was recorded on a daily or biweekly basis throughout the study as summarized in TABLE 13. A report of chemical usage was prepared on a biweekly basis to facilitate the calculation of average dosage rates for the two week periods.

Electric power and natural gas were metered at the plant entrance. The volume of fuel oil in storage was monitored. Fuel usage was not metered at the appropriate cost centers but it was readily estimated. The average line power consumption of each of the 20 largest motors was determined with the combination of a watthour meter (GE no 700X1365) and digital clocks to record time-on. The time-on was recorded for each of the 20 drives on a biweekly basis.

All materials, supplies, and services purchased for the Rosemount facility during the two year study were charged to the appropriate cost center/s.

TABLE 12. COST CENTERS

Cost Center	Title and Description
10	PRELIMINARY TREATMENT - plant inlet to discharge to wet well, includes flow measurement, bar screens, raw sampling pump
20	INFLUENT PUMPING STATION - wet well to discharge to clarifier
30	CLARIFICATION & PHOSPHORUS REMOVAL - inlet to clarifier to weir box on clarifier
40	FILTRATION-1 - weir box on clarifier to discharge to filtered water storage tank 1
50	CARBON ADSORPTION - filtered water storage tank 1 to weir box feeding 2nd stage filters
60	FILTRATION-2 - weir box to and including filtered water storage tank - 2
70	AMMONIA REMOVAL - ion exchange pumps to discharge from building
80	CHLORINATION - chlorine system for disinfection and odor control
90	EFFLUENT DISPOSAL - building to outfall
100	SLUDGE HANDLING & DISPOSAL - blowdown line (outside clarifier) to discharge from truck
110	CARBON REGENERATION - transfer and regeneration
120	CLINOPTILOLITE REGENERATION
130	AMMONIA RECOVERY SYSTEM - condenser and storage tank
140	AMMONIA DISPOSAL - disposition of stored ammonia solution
150	LABORATORY SERVICES - sampling, analysis, monitoring
160	BUILDING & GROUNDS - general systems including water, drain, compressed air, heating, ventilation, cleaning, etc.
170	INDIRECT SERVICES - nondirect labor, plant supervision, project supervision, clerical labor, data processing and other indirect services

TABLE 13. CHEMICAL INVENTORY SYSTEM

Chemical	Record
Ferric chloride	level measurement in day tank daily and storage tank, biweekly
Hydrated lime	measure concentration of each batch slurry elevated to storage tank, daily
Sulfuric acid	level measurement in storage tank, biweekly
Polymer	volume of liquid polymer added to dilution tank, daily
Nitrate	weight of sodium nitrate added to dilution tank, daily
Sodium chloride	number of 100 lb bags used, daily
Soda ash	weight used, daily
Sodium hydroxide	volume of 50% solution used, daily
Chlorine	rotometer settings on chlorinators, daily
Activated Carbon	number of 50 lb bags used, daily

VIII. WASTEWATER CHARACTERISTICS

During the two year reporting period the metered raw wastewater flow averaged 242,100 gpd. The sewered population in the City of Rosemount is estimated to have been approximately 3000 during the study. The recorded sewer services were as follows: residential - 800, institutional - 10, commercial - 35, and industrial - 2. The records of the Metropolitan Waste Control Commission indicate the industrial flows averaged 13,700 gpd and consisted primarily of sanitary sewage.

Flow variations throughout the day and from day to day were substantial as is typical for small sewer systems. The average dry weather influent hydrograph for the period October 27 through November 2, 1975 is presented in Figure 13. These data illustrate dry weather flow variations in excess of 4:1 (max:min). The variation of daily flows is characterized as follows: minimum - 85,000 gpd, maximum 642,000 gpd, and standard deviation - 51,000 gpd. The range of values is considerable because of infiltration and inflow during wet weather and because the main gate was periodically closed to allow for repairs. After the repairs were completed the wastewater accumulated in the interceptor was treated at higher than normal rates.

The chemical characteristics of the raw wastewater are summarized in TABLE 14 and Figure 14. The average values presented are typical of domestic sewage with the exception of the value of chloride and the relationship between ammonia and Kjeldahl nitrogen. The chloride and hardness concentrations in the Rosemount municipal water supply were determined on several occasions and found to be approximately 5 mg/l and 300 mg/l respectively.

The elevated chloride level of the raw wastewater was most likely due to the discharge of industrial and commercial wastes and the discharge of spent brine during the regeneration of individual water softeners of the ion exchange type. The relatively low value for organic N and corresponding high value for NH_3-N are most likely caused by the biochemical degradation of organic nitrogen compounds, and associated release of ammonia, in the intercepting sewer. Approximately 25,000 ft of pipe varying from 27 to 48 in. diameter are used to transport the wastewater from the edge of the sewered area to the treatment plant. The calculated velocity of flow is 1 fps for a flow rate of 0.25 mgd. During periods of low flow scour velocity was most likely not maintained and organic solids deposited.

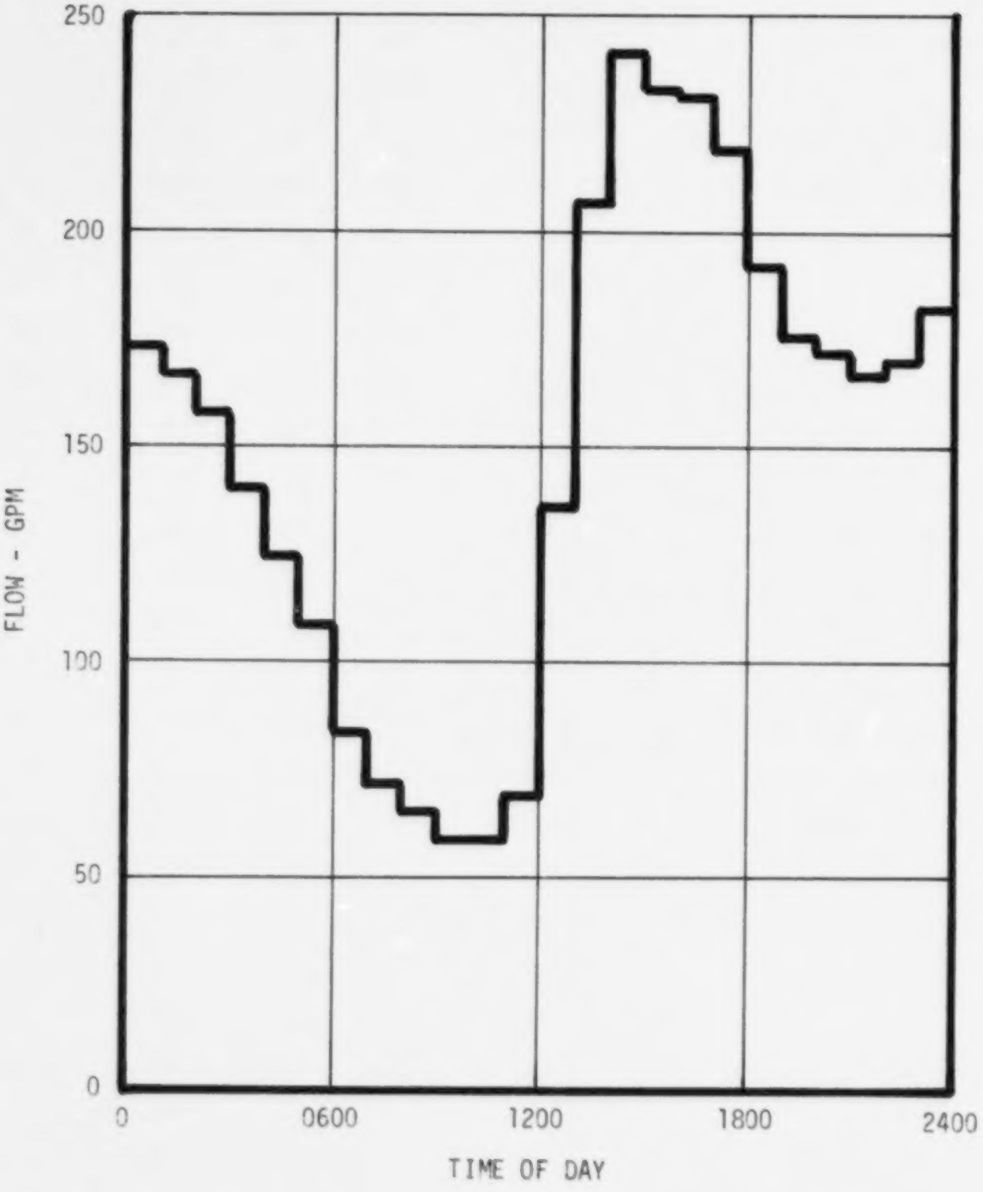


Figure 13. Typical dry weather influent hydrograph.

TABLE 14 RAW WASTEWATER CHARACTERISTICS

Parameter	Number of Analyses	Min	Max	Ave.	Std. Deviation
Temperature °F	703	44	66	55.5	4.5
pH	697	6.7	9.2	7.6	0.2
Alkalinity, mg/l $CaCO_3$	309	212	622	427	38
Hardness, mg/l $CaCO_3$	97	182	452	296	43
Calcium, mg/l $CaCO_3$	31	128	274	177	33
Chloride, mg/l	300	106	826	311	103
Total Solids, mg/l	93	996	2338	1426	272
Total Volatile Solids, mg/l	91	293	999	556	149
Suspended Solids, mg/l	319	29	1530	237	152
BOD - total, mg/l	316	71	999	211	84
BOD - soluble, mg/l	305	19	200	60	24
COD - total, mg/l	317	150	2435	505	234
COD - soluble, mg/l	301	51	350	143	36
TOC - total, mg/l	313	35	496	104	47
TOC - soluble, mg/l	232	11	200	42	19
Total Phosphorus, mg/l	299	4	52	11.7	3.7
Total Soluble Phosphorus, mg/l	283	2.6	25	8.5	2.3
Total Ortho Phosphorus, mg/l	305	3.1	26	9.2	2.4
Soluble Ortho Phosphorus, mg/l	302	2.4	24	8.2	2.3
Ammonia-nitrogen, mg/l	440	11	99	35	7
Kjeldahl-nitrogen, mg/l	258	19	270	42	17
Copper, mg/l	14	0.05	0.15	0.08	0.04
Nickel, mg/l	14	-	-	<0.05	-
Lead, mg/l	14	<0.05	0.20	-	-
Zinc, mg/l	14	0.17	0.57	0.27	0.10
Cadmium, mg/l	7	<0.02	<0.02	<0.02	-
Chromium, mg/l	14	<0.05	1.00	-	-
Iron	13	0.93	3.35	1.43	0.62

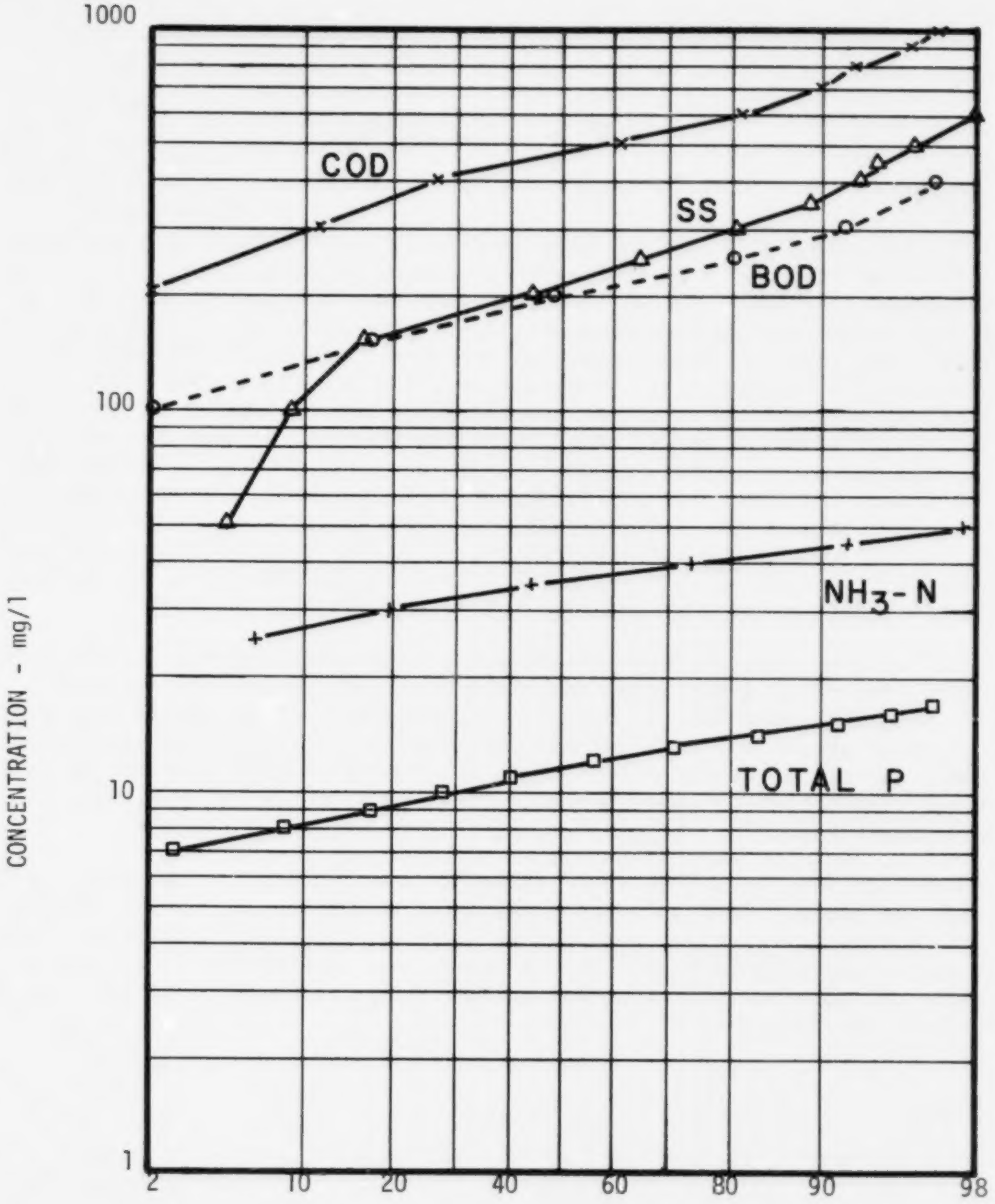


Figure 14. Frequency distributions for raw wastewater parameters.

IX. PLANT OPERATING CHRONOLOGY

A. GENERAL

The Rosemount AWTP was put into service in mid-November 1973 and since that time has been the only wastewater treatment facility serving the City of Rosemount. Although the on-line processes were all operational at that time, the carbon and zeolite regeneration systems were not available for service. The initial carbon regenerations were conducted in November 1974 and the zeolite regeneration system was put into service in May 1975. The original plant start up which was supervised by the contractor's representative lasted approximately 6 weeks. The contractor however, through his representative, retained the ultimate responsibility for plant operation through May 1975 although the plant was staffed by Metropolitan Waste Control Commission personnel. Although the facility was not formally accepted by the Metropolitan Waste Control Commission until November 4, 1975 the contractor was relieved of the direct responsibility for plant operations subsequent to the formal performance test which was conducted May 11 through May 15, 1975.

During the approximately 18 months the contractor was responsible for operating the plant a number of operating modes were used. For several months both treatment trains were used. During the entire period however the overall flow scheme was constant and consisted of: clarifier, dual media filters, one or two carbon columns, and dual media filters followed by chlorination. The operating modes for the clarifier/s and carbon columns are summarized in TABLE 15.

Prior to initiating the study it was planned to operate the plant utilizing six defined process configurations. Each configuration or mode of operation was to be used for an eight week period. The most successful mode was then to be used for approximately one year of operation. During the study however the schedule was modified to reduce the number of operating modes fo the five described in the following sections.

B. MODE I

The process configuration used during the eight week period June 7, 1975 through August 1, 1975 is illustrated in Figure 15. All the treatment units of train number 2 were utilized during this period. The clarifier was operated with a chemical feed combination of lime, ferric chloride and polymer. The carbon columns were operated in the series upflow mode. The zeolite columns were operated in the series downflow mode. The average raw sewage flow to the plant was approximately 0.22 mgd.

TABLE 15. CLARIFIER AND CARBON COLUMN STATUS

Month	Clarifier Chemicals	Carbon Column
1973 Nov.	$FeCl_3$ + Polymer	several comb.
Dec.		
1974 Jan.		
Feb.		2 col. series downflow
March		several comb.
April		2 col. series upflow
May		Col. 4,5 series upflow
June		
July	Lime + $FeCl_3$ + Polymer	
August		Col. 5,6 series upflow
Sept.		
Oct.		
Nov.	Lime + Polymer	Col. 4 downflow
Dec.	Lime + $FeCl_3$ + Polymer	
1975 Jan.		several combinations
Feb.		
March		2 trains, 2 columns series downflow
April		
May		

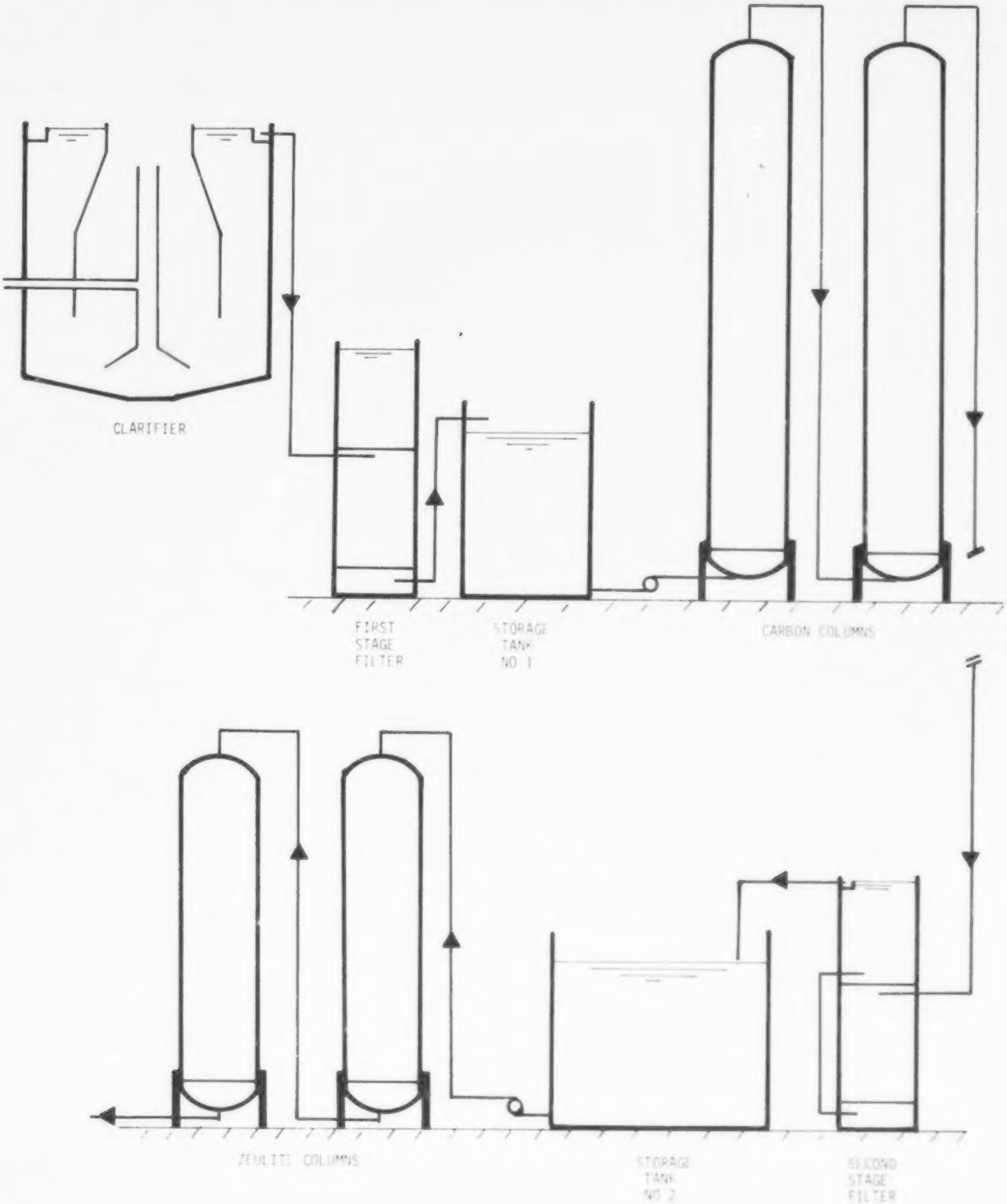

CLARIFIER
FIRST STAGE FILTER
STORAGE TANK NO 1
CARBON COLUMNS
ZEOLITE COLUMNS
STORAGE TANK NO 2
SECOND STAGE FILTER

Figure 15. Flow scheme during Mode I operation.

The operating conditions of the clarifier are summarized in TABLE 16. The lime and ferric chloride dosages were relatively constant throughout the eight week period. The polymer dose averaged 1.5 mg/l during the first four weeks and 4.2 mg/l during the second four weeks. The increased polymer dose was required by the operators to control the sludge bed level.

First stage filters number 3 and 4 and second stage filters number 7 and 8 were in service during this reporting period. The filters were taken out of service for backwashing only.

TABLE 16. CLARIFIER OPERATING CONDITIONS - MODE I

Average flows - gpd	
raw	222,000
clarifier	277,000
Mixing zone pH	
average of 1290 hourly values	10.5
Effluent pH	
set point	7.5
average observed	7.5
Average chemical additions	
lime - mg/l CaO	424
ferric chloride - mg/l Fe	14.1
polymer - mg/l Nalco 677	2.9
sulfuric acid - mg/l SO_4-S	51

The backwashing frequencies were as follows:

filter no.	no backwash cycles
3	77
4	77
7	59
8	58

The backwash water flow amounted to approximately 14,500 gpd.

Carbon columns 4 and 5 were operated in the series upflow mode. Column number 4 was regenerated and put into service on March 20, 1975. It was used as a primary column in the downflow mode for 21 days and as a primary upflow column for 39 days prior to June 7, 1975. Column under 5 was regenerated and put into service on May 5, 1975. It was used as a secondary upflow column for 33 days prior to June 7, 1975. The carbon regeneration data are as follows:

		Apparent Density		I_2 Number	
Date	Column	Spent	Regen	Spent	Regen
3/20/75	4	0.53	0.52	710	795
4/28/75	5	0.49	0.46	850	1020

The regeneration of column 4 was not complete because of scrubber, induced draft fan, and burner outages during the run.

A control system was provided to recirculate flow around the carbon columns tomaintain the upflow rate equal to, or greater than, a manually entered set point flow. In the automatic mode the control valve cycled. In order to avoid mechanical damage to the piping system the recycle system was operated in the manual mode. However the system received little operator attention and the recycle valve remained open during periods when the forward flow was in excess of the minimum flow set point. The flow through the carbon columns was thus diluted and the contact time needlessly reduced. Recycle was terminated on July 28, 1975.

The carbon columns were backwashed at approximately 5 day intervals and at a rate of 12 gpm/sq ft. Columns 4 and 5 were backwashed 11 and 10 times respectively during the reporting period. The backwash water requirement was approximately 3400 gpd.

The ion exchange columns were in operation for 22 days during the reporting period. It was found that because of the constraints of time and limited storage capacity for spent backwash water only one complete regeneration-strip cycle could be routinely accomplished per day. The regenerations were conducted with a 1N sodium solution at a pH of 11.0. The regeneration sequence utilized is summarized in TABLE 17. Approximately 17,700 gal were recycled to the headworks via the backwash pit during each regeneration cycle. Because the columns were in service only 22 days the recycle flow averaged only 6950 gpd.

C. MODE II

The mode II configuration illustrated in Figure 16 was used a total of 34 days for the periods August 2-21, September 5-12 and September 25-30, 1975. Treatment train number 2 was used with the first stage filters out of service - bypassed. The zeolite columns were out of service during the entire period

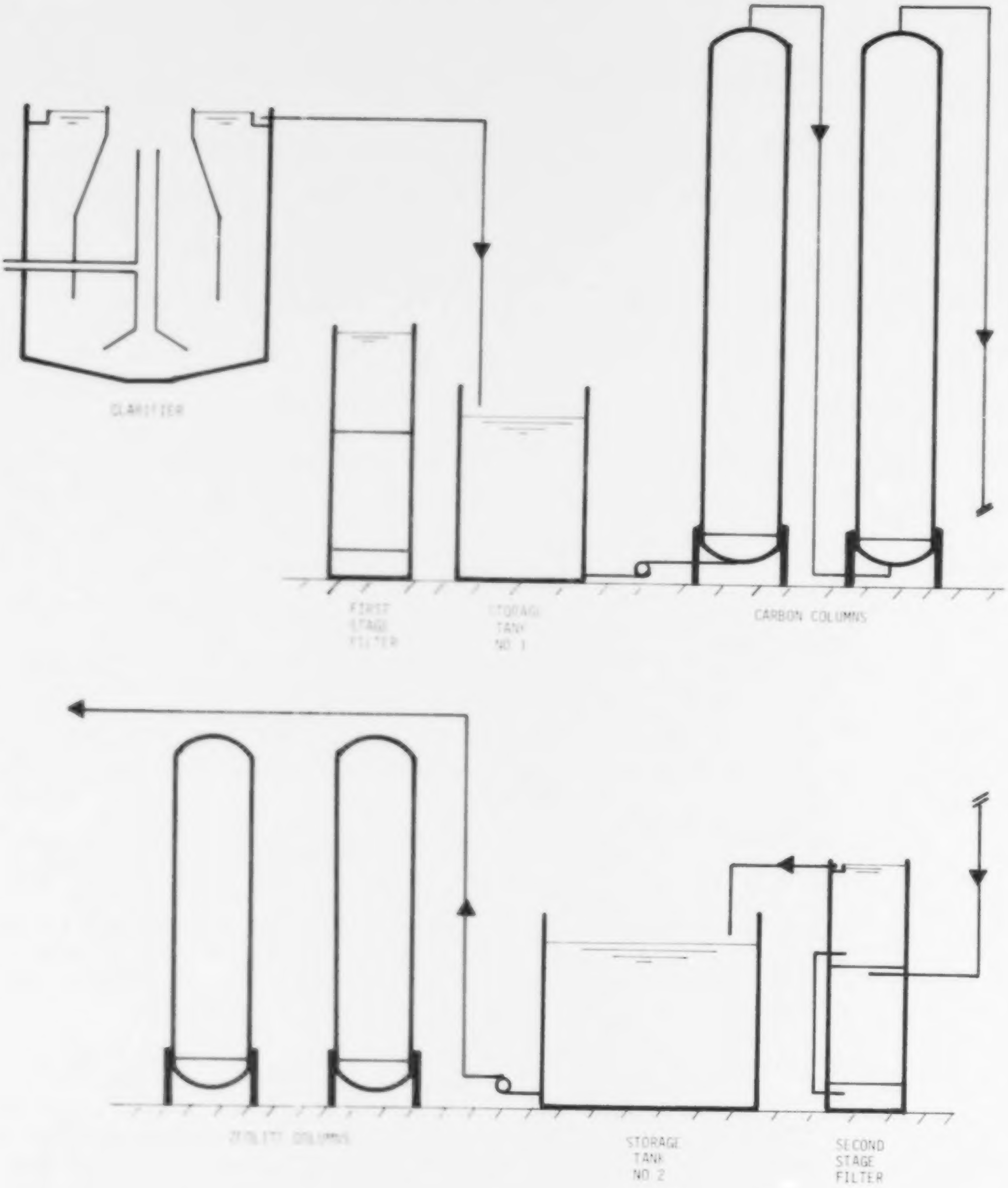

CLARIFIER
FIRST STAGE FILTER
STORAGE TANK NO 1
CARBON COLUMNS
ZEOLITE COLUMNS
STORAGE TANK NO 2
SECOND STAGE FILTER

Figure 16. Flow scheme during Mode II operation.

because of problems with the regeneration system.

TABLE 17. ZEOLITE REGENERATION SEQUENCE - MODE I

Function	Control
Column backwash - 420 gpm	10 min - timer
Settle bed	5 min - timer
Brine upflow - water to waste	conductivity control
Brine upflow to recovery - 100 gpm	low level in brine tank
Air displacement of brine to recovery	120 min - timer
Refill upflow - 100 gpm	45 min - timer
Settle bed	5 min - timer
Rinse to waste - 150 gpm	60 min - timer

The average clarifier conditions are summarized in TABLE 18. The lime dose was relatively constant but the ferric chloride and polymer concentrations varied substantially as illustrated in TABLE 19. The polymer dose was reduced intentionally to eliminate the possibility of polymer discharge from the clarifier. At that time polymer coating or blinding of the activated carbon granules was considered a possibility.

The drastic reduction in ferric chloride dose occurred when the feed pump was adjusted to give a dose of 7 mg/l as Fe. It was found however that the pump calibration was not accurate in the extreme low range. At that time the feeders pumped directly from the ferric chloride storage tank to the clarifier. Because the tank is 8 feet in diameter, it was difficult to determine consumption for periods of less than 10-14 days. Two 55 gallon day tanks were put into service to eliminate this problem - Figure 7.

The first stage filters (number 3 and 4) were taken out of service at 2400 hr August 1, 1975. After about one week of operation, the backwash requirements of the second stage filters increased substantially as illustrated in TABLE 20. On August 13 both filters were chlorinated prior to backwash to eliminate the problem of rapid head loss increases. This procedure had been used successfully during the previous year to remove the scum formed on the surface of the second stage filters. During the week of August 14-21 the head box feeding filters 7 and 8 overflowed on numerous occasions. These overflows discharged to the backwash pit and substantially increased the plant recycle flow.

TABLE 18. CLARIFIER OPERATING CONDITIONS - MODE II

Average flows - gpd	
raw	226,000
clarifier	285,000
Mixing zone pH	
average of 782 hourly values	10.6
Effluent pH	
set point	7.5
average observed	7.6
Average chemical additions	
lime - mg/l CaO	474
ferric chloride - mg/l Fe	10.5
polymer - mg/l Nalco 677	2.8
sulfuric acid - mg/l SO_4-S	56

TABLE 19. CLARIFIER CHEMICAL FEED - MODE II

Period	$FeCl_3$-mg/l Fe	Polymer-mg/l Nalco 677
August 2-21, 1975	14.5	4.1
September 5-12, 1975	2.5	1.0
September 25-30, 1975	10.5	0.4

TABLE 20. FILTER BACKWASH FREQUENCY - MODE II

Date	Filter Number 7	 8
August 2	2	2
3	1	1
4	1	1
5	1	1
6	2	2
7	2	2
8	2	1
9	5	4
10	3	2
11	2	2
12	3	2
13	6	5
14	3	2
15	3	3
16	3	3
17	2	2
18	3	3
19	2	1
20	2	2
21	3	3
September 5	2	2
6	1	1
7	1	1
8	3	3
9	2	3
10	2	3
11	3	3
12	1	3
25	1	3
26	5	4
27	1	1
28	2	2
29	1	1
30	1	2
Total	77	76

The first stage filters (number 3 and 4) were put back into service on August 22. On August 27, filter number 8 was drained down and the media surface exposed. The surface mat had a consistency of wet clay and could easily be formed by hand. The surface mat, approximately 1/2 inch deep, was raked off and the filter put back into service.

On September 3, the headbox serving the second stage filters was examined. It was observed that when one filter was being backwashed or had reached its hydraulic capacity and flooded its headbox inlet the headbox would overflow before the second filter reached its hydraulic capacity. The headbox design was modified to eliminate this problem.

Based on suggestions from the plant operators and laboratory testing, it was found that the filter media cleansing was expedited by a presoak in a solution of 0.2N sodium chloride. This cleaning technique was used on filters 7 and 8 on September 2. The media was soaked in the salt solution for approximately 30 minutes. The media was then air scoured for 15-20 minutes prior to normal backwash.

The first stage filters were taken out of service early on September 5. The backwash requirements for the second stage filters again gradually increased. On September 12 during an attempt to treat the carbon columns with a salt solution a considerable quantity of scum, grease and other debris was released from the walls of filtered water storage tank number 1. All of the material had been deposited below the water line. The tank was taken out of service and thoroughly cleaned.

A carbon regeneration was initiated on September 13. The spent scrubber water from the carbon furnace discharges to the clarifier effluent line. With the first stage filters out of service, any solids in the spent scrubber water would have been deposited directly in the clean storage tank. For this reason the first stage filters were put back into operation on September 13. When the regeneration was terminated the first stage filters were again taken out of service - September 25.

On October 1, storage tank number 1 was drawn down and inspected. Grease deposits and scum were observed on the water surface and tank walls. The first stage filters were put back into service to reduce or eliminate the transport of more of this material to the storage tank and eventually to the carbon columns. The attempt to operate the plant with the first stage filters out of service was terminated.

For the reporting period the filter backwash requirements averaged approximately 18,000 gpd. The backwash requirement was not substantially reduced during the period the first stage filters were out of service.

Carbon columns 4 and 5 were operated in the upflow series mode until August 11. At that time column 4 was taken out of service, column 5 was moved to the primary position and column 6, which was freshly regenerated, was brought into service in the secondary position - again in the upflow mode. For the reporting period, the backwash requirement averaged approximately 9500 gpd.

D. MODE III

The process configuration during mode III was identical to that of mode I as illustrated in Figure 15. Data were collected for the 78 day period August 22 - September 4, September 13-24, and October 1 - November 21, 1975.

The average operating conditions for the clarifier are summarized in TABLE 21. For the 7 day period August 23-29 no lime was available for use in the clarifier because of a late delivery. During the interim period $FeCl_3$ was used as the prime coagulant at an average dose of 54 mg/l Fe along with polymer at an average dose of 4.1 mg/l. All polymer use was terminated on November 16, 1975.

Both the first and second stage filters were in service (number 3, 4, 7, 8) during the reporting period. The backwash requirement averaged approximately 23,200 gpd. This was significantly higher than the backwash requirements in Modes I and II. During the 4 day period August 26-29 the head loss buildup in the first stage filters was more rapid than usual and frequent backwash was required. During this same period difficulty was encountered in the laboratory in the suspended solids analysis. It was not possible to filter more than 50 ml of raw sewage through an 11 cm glass fiber filter although the solids concentrations were normal. Under normal conditions 250 ml was filtered easily. On August 30 and 31 the two first stage filters were backwashed a total of 24 times, once with a salt solution. Backwash requirements for the primary filters were normal after August 31. After those periods during which the primary filters were out of service the secondary filters required frequent backwash and special cleansing to remove the scum and debris collected.

Carbon columns 5 and 6 were operated in the series upflow mode for the entire reporting period.

The ion exchange columns were in service for 15 days near the end of the reporting period. Columns 5 and 6 were used on alternate days for 24 hour periods. The effects of multiple regenerations and regeneration at pH values greater than 11 on zeolite exchange capacity were determined.

E. MODE IV

The general process configuration illustrated in Figure 17 was utilized for a period of 420 days, November 22, 1975 - January 14, 1977. During the first 9 days of the period only one carbon column was in service in the downflow mode. Two columns in the series downflow mode were used for the following 411 days. The zeolite columns were used for a total of 123 days with one or two columns in service.

The clarifier conditions are summarized in TABLE 22. For the first 15 days of mode IV the internal pH of the clarifier was maintained near 10.9. For the periods December 6, 1975 through 1200 hr June 16, 1976 and 1200 hr July 9, 1976 through January 14, 1977 lime was fed at a reduced rate to maintain

TABLE 21. CLARIFIER OPERATING CONDITIONS - MODE III

Average flows - gpd	
raw	209,000
clarifier	292,000
Mixing zone pH	
average of 1753 hourly values*	10.5
Effluent pH	
set point *	7.5
average observed *	7.5
Average chemical additions *	
lime - mg/l CaO	492
ferric chloride - mg/l Fe	9.1
polymer - mg/l Nalco 677 - terminated November 16, 1975	0.95
Sulfuric acid - mg/l SO_4-S	64

* excluding data for August 23-29, 1975

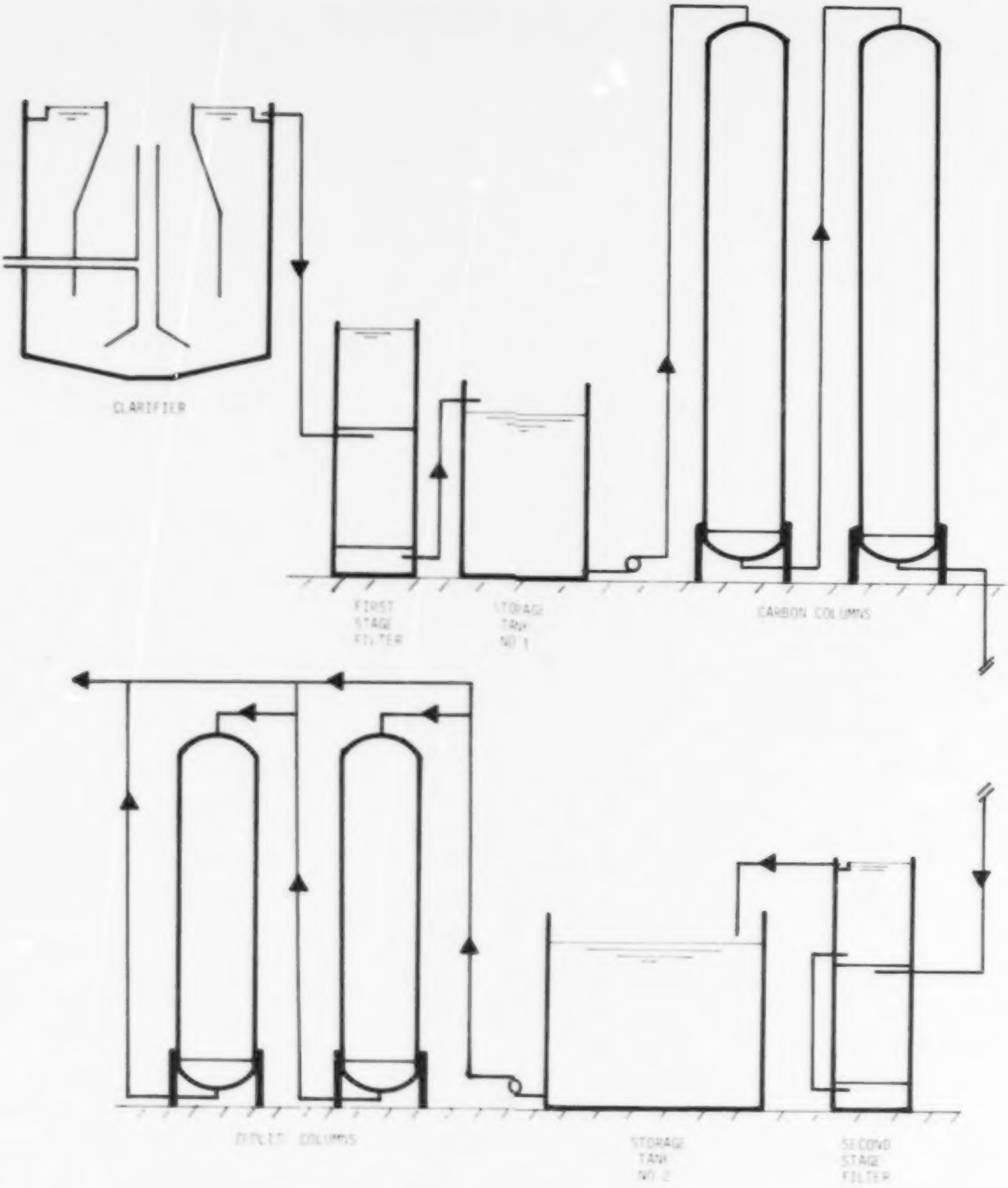

CLARIFIER
FIRST STAGE FILTER
STORAGE TANK NO 1
CARBON COLUMNS
ZEOLITE COLUMNS
STORAGE TANK NO 2
SECOND STAGE FILTER

Figure 17. Flow scheme during Mode IV operation.

TABLE 22. CLARIFIER OPERATING CONDITIONS - MODE IV

Average flows - gpd	
raw	238,600
clarifier	293,000
Mixing Zone pH	
Nov. 21 - Dec. 5, 1975 - Period I average of 336 hourly values	10.9
Dec. 6, 1975 - June 16, 1976, 1200 hr) July 9, 1976, 1200 hr - Jan. 14, 1977) Period II average of 8714 hourly values	9.6
Effluent pH	
set point, Periods I & II	6.5 to 7.5
average observed Periods I & II	7.5
June 16, 1976, 1200 hr - July 9, 1976, 1200 hr - Period III average observed	7.5
Average chemical additions	
lime - mg/l CaO	
Period I	467
Period II	338
Period III	0
ferric chloride - mg/l Fe	
Periods I & II	18.5
Periods III	54
polymer - mg/l Nalco 677	
Periods I & II	0
Periods III	5.7
sulfuric acid - mg/l SO_4-S	
Period I	76
Period II	63

TABLE 23. CARBON COLUMN STATUS - MODE IV

Dates		Treatment Train	Carbon Columns in service
From	To		
11/22/75	12/30/75	2	4
12/31/75	2/17/76	2	4,5 series
2/17/76	2/26/76	1 & 2	2,3 series & 4,5 series
2/26/76	3/1/76	1	2,3 series
3/2/76	5/17/76	2	4,5 series
5/18/76	8/24/76	2	5,6 series
8/25/76	11/22/76	1	1,2 series
11/23/76	12/9/76	1	2,3 series
12/10/76	12/17/76	2	5,6 series
12/18/76	1/14/77	1	3,1 series

a pH of approximately 9.5. For the 23 day period, 1200 hr June 16 through 1200 hr July 9, 1976, ferric chloride was used as the primary coagulant along with a substantial quantity of polymer. This change was necessitated by a prolonged labor dispute involving the regular lime supplier.

The status of the treatment trains and carbon columns are presented in TABLE 23. Because of maintenance requirements both treatment trains were used during this reporting period. However both trains were used simultaneously for only 9 days during a period of extremely high raw wastewater flow which was caused by inflow/infiltration during an early spring storm. The treatment trains were alternated to accomplish required maintenance such as cleaning rags from the clarifier mixing impeller and replacing valves on the filters and carbon columns. Both the primary and secondary dual media filters were used throughout the period.

A number of combinations of brine pH, brine volume and sodium concentration were used for zeolite regeneration. For this reason the changes dealing with the zeolite regeneration process are presented in a separate section.

F. MODE V

The process configuration illustrated in Figure 18 was utilized for 143 days, January 15, 1977 to June 6, 1977. Process train No. 1 was used throughout the period. For a period of 5 days in March both treatment trains were used to accommodate the high flows associated with snow melt. The second stage filters were not in service during this operating mode.

The clarifier conditions are summarized in TABLE 24 and the status of the carbon columns is presented in TABLE 25. As in mode IV several zeolite regeneration procedures were used and are summarized in the following section.

G. AMMONIA REMOVAL SYSTEM

Although the treatment plant was put into service in November of 1973 the ammonia removal system was not completed and tested until May of 1975. For this reason the plant personnel did not have a great deal of operating experience with this system when the demonstration project was initiated. During the period June 7, 1975 through August 12, 1977 the zeolite columns were in service a total of 277 days and a total of nine operating modes were evaluated as summarized in TABLE 26. For the period prior to March 1977 the ammonia removal system was in service on only a sporadic basis because of a series of problems related to design, construction and installation and because of the energy crisis which prevailed during the winter of 1976-77.

During period number 1 - TABLE 26 - the system was operated with two zeolite columns in series. At midnight each day the primary column was taken out of service for regeneration, the secondary column was moved to the primary position and a freshly regenerated column was put into service in the secondary position. The changes in column status were made on a time basis so as not to

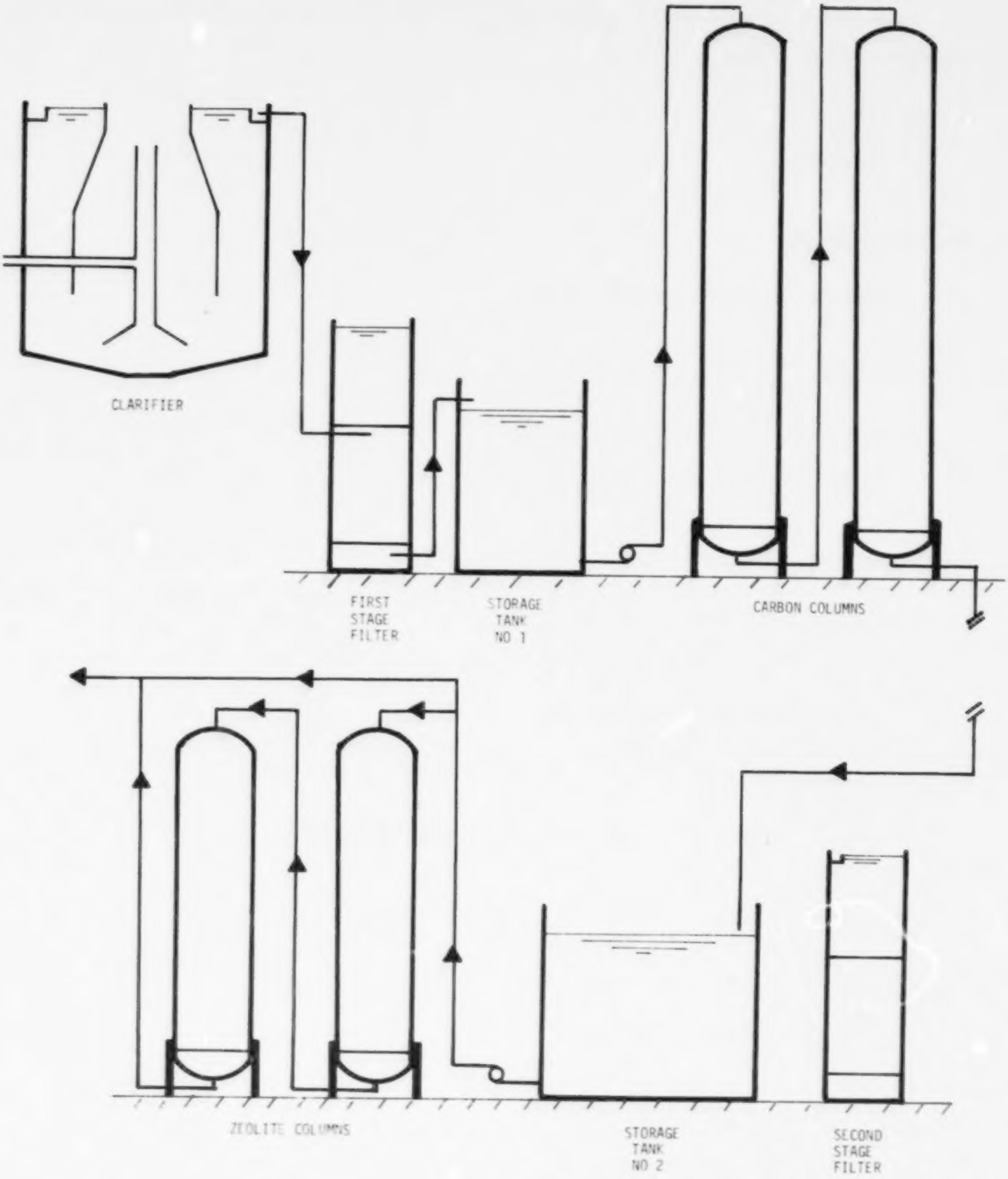

CLARIFIER
FIRST STAGE FILTER
STORAGE TANK NO 1
CARBON COLUMNS
ZEOLITE COLUMNS
STORAGE TANK NO 2
SECOND STAGE FILTER

Figure 18. Flow scheme during Mode V operation.

TABLE 24. CLARIFIER OPERATING CONDITIONS - MODE V

Average flows - gpd	
raw	280,900
clarifier	327,600
Mixing zone pH	
average of 3225 hourly values	9.75
Effluent pH	
set point	6.5 to 7.5
averaged observed	7.3
Averaged chemical additions	
lime - mg/l CaO	327
ferric chloride - mg/l Fe	19.6
sulfuric acid - mg/l SO_4-S	52

TABLE 25. CARBON COLUMN STATUS - MODE V

Dates		Treatment Train	Carbon Columns in Series
From	To		
1/15/77	2/28/77	1	3,1 series
3/1/77	3/8/77	1	1,2 series
3/9/77	3/13/77	1 & 2	1,2 series, 5,6 series
3/14/77	6/6/77	1	1,2 series

TABLE 26. SERVICE AND REGENERATION CONDITIONS FOR AMMONIA REMOVAL SYSTEM

Operating Scheme No.	Period	Days of Operation	Columns in Service	Regeneration Conditions			Brine Recovery Conditions	
	Dates			pH	Na-N*	Brine-BV**	lb steam / gal brine	Complete Regenerations per Strip Cycle
1	June 7-July 20, 1975	23	2, series	11	1.0	4.6	1	1
2	November 6-December 9, 1975	28	1	12.5	1.0	4.6	1	1
3	December 10-December 25, 1975	13	1	13	1.0	4.6	1	1
4	March 2-June 14, 1976	32	2, series	13	0.5	4.6	1	1
5	June 16-June 30, 1976	14	2, series	12	0.5	4.6	1	1
6	July 3, 1976-April 26, 1977	84	2, series	12	0.5	8.2	1	1
7	April 30-June 9, 1977	35	2, series	12	0.5	14	1	1
8	June 12-July 8, 1977	27	2, series	12	0.5	14	0.75	1
9	July 23-August 12, 1977	21	2, series	12	0.5	14	0.75	2

* 1.0 Normal Solution = 23 g/l Na

** BV = Empty Bed Volume = 2250 gal.

disrupt the scheduling for the remainder of the plant. Because of the severe raw flow variations (Figure 13) a serious attempt was made to return all recycle flows to the headworks during periods of low raw flow and thus minimize fluctuations in hydraulic loading throughout the plant. This required that all recycle flows be generated on a scheduled basis because of the limited storage capacity for spent backwash water.

The teflon lining of the brine transfer pump casing failed on July 10, 1975. This failure allowed hot brine to leak from a weep hole at a rate of approximately 5 gpm. The regeneration system was out of service for approximately 3 weeks while the pump was repaired. After 23 days of operation it was apparent that the zeolite was not being thoroughly regenerated because the ammonia removal efficiency was well below design values. It was also apparent that the steam stripping operation was not functioning properly because the observed condensate volumes were approximately three times the design volume.

After several test runs it was determined that the stripper product level control system had malfunctioned and allowed the stripping column to fill with brine and overflow to the ammonia receiver. The control system piping was modified, all instruments calibrated and a sight glass installed in the column to allow visual determination of the liquid level if required. The regeneration system was put back into service in the 4th week of September 1975.

Several plant scale experiments were conducted during October 1975 to identify the ammonia elution characteristics as a function of brine volume and pH. As a result of these tests the regenerations during period number 2 were conducted at pH 12.5. For this period of 28 days zeolite columns number 5 and 6 were used on alternate days. Only one column was used at a time because the sampling system had not yet been modified to allow collection of the AEC_1 sample and it was desirable to identify ammonia removals per column.

On December 11, 1975 the regeneration procedure was again modified and for a period of 13 days the regenerations were conducted at pH 13. The ammonia removal system was taken out of service at midnight December 25, 1975 because substantial leaks had developed in both heat exchangers - Figure 9. The heat exchangers were dismantled and the 170 plates cleaned individually. All of the gaskets were replaced and several of the plates which had ruptured were replaced. After the heat exchangers were repaired a system was constructed to allow periodic in situ cleaning with dilute nitric acid.

The zeolite columns were put back into service on March 2, 1976 with two columns operated in series as in period number 1. The regenerations were conducted at pH 13 with a 0.5N NaCl solution. Several tests were made to determine the need for heated brine storage. Based on the results obtained the heat input (steam-heat exchanger) was terminated and both of the heat exchangers (Figure 9) were removed from the system.

Because of the extremely high cost for caustic and rinsing requirements the regeneration pH was brought down to 12. The system was operated in this manner for 14 days - period number 5. Starting in July 1976 the regeneration procedure was modified to yield a 78% increase in brine volume for regeneration.

The initial 4.6 bed volumes, BV, were passed through the zeolite following the normal procedure, however, the spent brine was softened and passed through the zeolite a second time. The pH was adjusted as required prior to the second pass. During periods number 5 and number 6 it was observed that the ammonia removal efficiency in the stripping column had decreased because of the lower brine inlet temperature. One of the original heat exchangers was installed to preheat the brine to the stripper and in turn cool the brine discharged from the stripper. This modification was completed August 10, 1976. For the period August 15 to October 21, 1976 the zeolite columns were out of service to allow for repair of all the system valves some of which did not seal properly. The system was out of service for the period December 2, 1976 through March 21, 1977 because of the failure of the teflon lining of the brine transfer pump casing and because of the energy crisis which existed in the State of Minnesota during the months of January and February 1977.

In late April 1977 the regeneration procedure was modified to increase the brine throughput to 14 BV during each regeneration cycle. The steam rate used for ammonia stripping was reduced by 25% in mid-June 1977. The final modification during the reporting period was made on July 23, 1977 and consisted of eliminating 50% of the stripping cycles by stripping on alternate days. The data obtained during the period June 5 through August 12, 1977, periods number 8 and number 9 are included because they are significant in terms of system evaluation.

X. OVERALL PLANT PERFORMANCE

A. PRIOR TO DEMONSTRATION PROJECT

Despite the frequent discontinuities associated with plant startup and shakedown the Rosemount AWTP produced a high, although somewhat variable, quality effluent during its first year and one half of operation. The data summary presented in TABLE 27 illustrate this fact. Some of the variability in treatment efficiencies during the first year of operation was also due, no doubt, to the fact that the operating personnel, although experienced in conventional wastewater treatment, had no experience with the unit processes utilized at Rosemount. The effluent suspended solids data demonstrate that the experience of the plant operators, although not readily quantified, is an important parameter when making realistic estimates for future plant performance.

Several combinations of ferric chloride, lime and polymer were used to remove phosphorus and aid solids removal. The treatment goal was an effluent total phosphorus concentration of 1 mg/l. Experience in laboratory jar tests and in the clarifier indicated that any combination of ferric chloride, lime and polymer that effected the required phosphorus removal would also yield high removals of suspended solids. The operator attention required to assure high solids removals was found to vary substantially. During the 7 1/2 month period, November 1973 - June 1974, that ferric chloride was used as the primary coagulant (average approximately 45 mg/l Fe) the clarifier was difficult to control and subject to upset on a daily basis because of the unsteady hydraulic loading and the characteristics of the ferric hydroxide floc. In order to prevent significant floc carry over to the filters the operators would have to increase the polymer dosage during periods of increasing and high flow. On at least two occasions the sludge bed discharged to the launder and plugged the first stage dual media filters. On both occasions the filters were out of service for several hours because the standard backwash procedure did not provide sufficient cleansing. Based on this experience, the operators would often dump several pounds of polymer directly into the raw sewage wet well in order to control a rising sludge blanket.

During the 11 month period that lime was used, along with ferric chloride and polymer or with polymer alone, the problem of floc carry over was greatly reduced. The lime-ferric chloride combinations varied in the pH range of 9 to 10.5 and from 4.5 to 45 mg/l Fe. For the 30 day period November 7 through December 6, 1974 the clarifier was maintained at a pH of approximately 11.4 and polymer used as a coagulant aid. Solids removal in the clarifier and through the plant was consistently high; however, it was not always possible to maintain the effluent total phosphorus below 1 mg/l.

TABLE 27. PLANT PERFORMANCE DATA FOR PERIOD PRIOR TO DEMONSTRATION PROJECT

Month	Average Flow - mgd	Coagulant	Carbon Columns	Concentration - mg/l BOD Inf	BOD Eff	SS Inf	SS Eff	Total P Inf	Total P Eff	NH_3-N Inf	NH_3-N Eff
January, 1974	0.182	$FeCl_3$ + Polymer	Several Combinations	113	19	117	24	-	0.5	-	37
February	0.194			167	21	122	15	-	0.7	-	44
March	0.234		2 trains 2 columns series downflow	299	13	265	6	-	0.3	-	31
April	0.229			282	9	286	5	-	-	-	-
May	0.206			216	17	237	12	-	-	-	-
June	0.214			292	22	341	7	-	-	-	-
July	0.205	Lime + $FeCl_3$ + Polymer	2 columns, parallel, down	223	30	239	9	-	1.1	26	23
August	0.200		2 columns series upflow	216	31	216	11	11.1	1.9	29	25
September	0.217			246	37	214	7	14.0	2.2	35	28
October	0.198			299	27	281	6	13.8	1.0	42	32
November	0.189	Lime + Polymer	1 column, upflow	264	14	238	5	13.0	0.9	37	35
December	0.205	Lime + $FeCl_3$ + Polymer	several combinations	301	34	209	4	14.4	1.1	45	38
January, 1975	0.222			264	19	253	3	13.3	1.2	38	35
February	0.219		2 columns, series, down	263	19	225	3	12.9	1.0	39	33
March	0.258		several combinations	195	16	235	2	11.9	1.0	35	29
April	0.258		2 columns series upflow	180	10	170	3	11.5	0.8	40	28
May	0.234			197	5	249	2	13.2	0.4	41	25
Average	0.216	-	-	230	20	229	7	13.0	1.0	37	32

The effluent BOD values demonstrated a high degree of variability because of the number of operating modes that were tried without regard for the past history of the carbon columns. It was necessary to alternate between upflow and downflow operation to diagnose more completely problems with the upflow distributors. Because of problems with the carbon furnace the first successful carbon regeneration was not made until November of 1974 and, thus, partially spent carbon was utilized in all columns for most of 1974.

No significant ammonia removals were accomplished until April and May 1975 when the contractor started to test the zeolite columns and the regeneration system.

B. DURING DEMONSTRATION PROJECT

1. General

Throughout the period of the demonstration project the Rosemount AWTP was operating under the authorization of NPDES permit MN0025488 which was issued by the Minnesota Pollution Control Agency on June 2, 1975. The permit conditions are summarized in TABLE 28.

TABLE 28. DISCHARGE PERMIT CONDITIONS

Effluent Characteristic	7 day average	30 day average Normal	30 day average Experimental	Notes
BOD - mg/l	-	15	25	arithmetic mean
SS - mg/l	-	15	30	arithmetic mean
Fecal Coliform per 100 ml	400	200	200	MPN
pH	range of 6.5 to 8.5 for instantaneous values, not subject to averaging.			
	no discharge of floating solids or visible foam in other than trace amounts.			
	discharge shall not contain oil or other substances in amounts sufficient to create a visible color film on the surface of the receiving waters.			

The two sets of BOD and SS standards were provided to accommodate the experimentation during the demonstration project. In terms of effluent quality the plant operations staff had three goals when the project was initiated as follows:

(a) to meet the discharge permit conditions;

(b) to maintain the monthly average total phosphorus concentration of 1.0 mg/l;

(c) to maintain the monthly average ammonia nitrogen concentration of 1.0 mg/l.

The overall plant performance during the two year period is characterized by the data summaries presented in Figure 19 and TABLE 29. Individual BOD values were less than or equal to 15 mg/l and 25 mg/l 54 and 89 percent of the time respectively. Individual suspended solids values were less than or equal to 15 mg/l 99.7 percent of the time. In like manner the effluent total phosphorus goal of 1 mg/l was exceeded only approximately 17 percent of the time.

The BOD removal efficiency could have been improved by regenerating carbon at more frequent intervals; however, one of the purposes of the evaluation was to document the effect of service time and/or organic loading on carbon performance.

After several months of operation it became obvious that the goal of 1 mg/l NH_3-N was not attainable. The new endeavor was to keep the ammonia removal system in service as much as possible and to optimize removal efficiencies with the existing facilities.

Most of the laboratory analytical data that relate to process performance are summarized in TABLE 29. Although these data are difficult to interpret because the sampling frequencies were not identical throughout the two year period several observations can be made as follows:

[a] Overall removal efficiencies averaged approximately 93%, 99%, and 94% for BOD, suspended solids and total phosphorus respectively.

[b] Based on average raw and clarifier flows of 0.242 mgd and 0.298 mgd respectively the removals accomplished in the clarifier averaged approximately 75%, 90%, and 90% for BOD, suspended solids and total phosphorus respectively.

[c] With the exception of iron all measurable amounts of the heavy metals analyzed were removed in the clarifier.

[d] The two stages of filtration accounted for approximately 5% of the BOD and suspended solids removal.

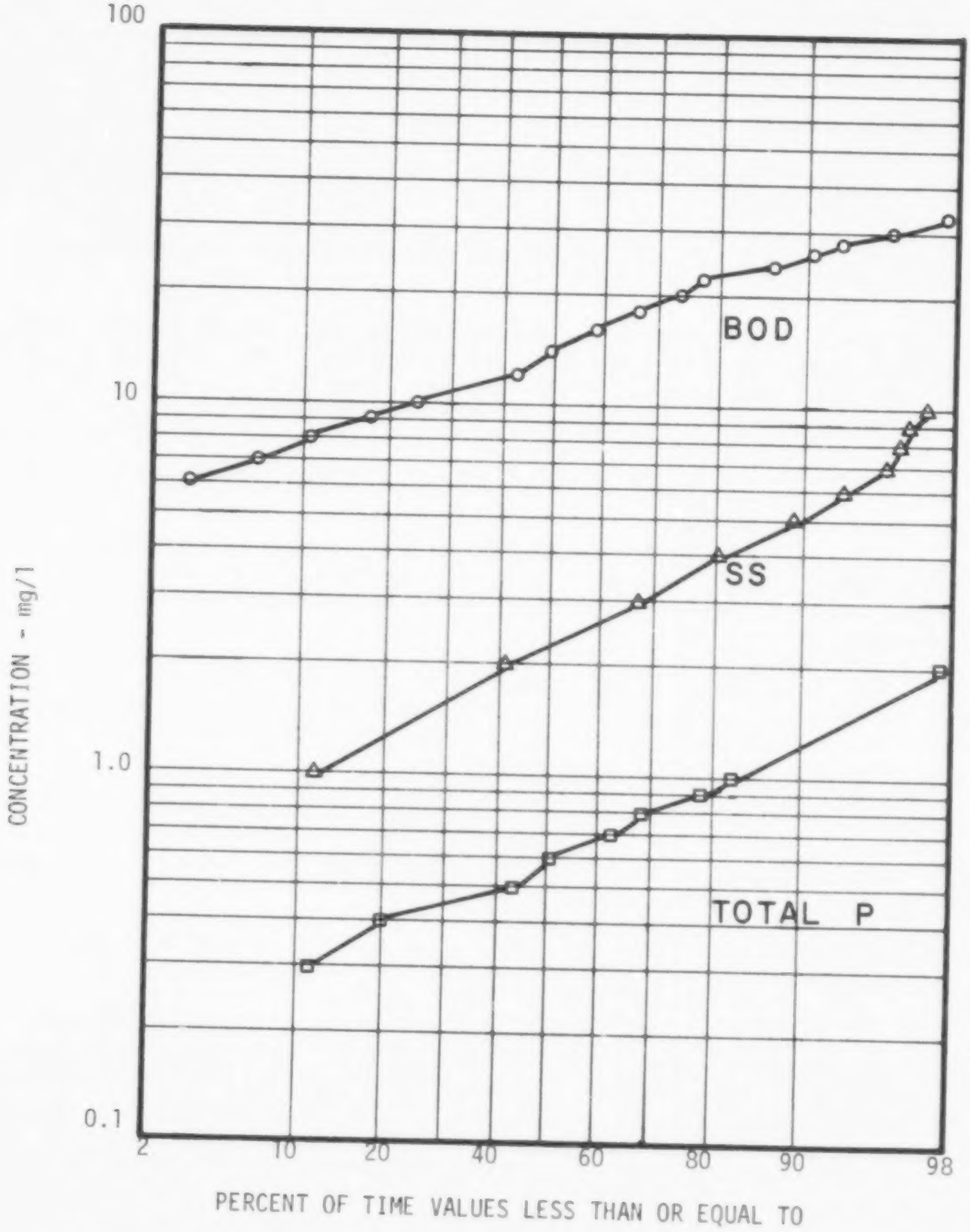


Figure 19. Frequency distributions for effluent paramaters during demonstration project.

TABLE 29. SUMMARY OF ANALYTICAL DATA

	Average Concentrations										
Parameter	RAW	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	211	175	41	31	18	18	15	-	-	-	15
BOD - soluble, mg/l	60	47	24	26	14	12	-	-	-	-	14
COD - total, mg/l	505	439	91	74	48	38	36	38	-	-	40
COD - soluble, mg/l	143	118	62	58	40	26	-	-	-	-	33
TOC - total, mg/l	104	89	24	21	13	10	9	10	-	9	11
TOC - soluble, mg/l	42	33	18	17	12	8	-	-	-	-	10
Suspended Solids, mg/l	237	255	19	10	5	7	6	2	-	3	3
Turbidity, NTU	-	-	9	7	4	7	8	6	3	5	7
Settleable Solids, ml/l		-	-	-	-	-	-	-	-	-	0
Temperature, °F	55.5	-	-	-	-	-	-	-	-	-	59.8
pH	7.6	7.8	7.5	7.5	7.5	7.5	7.6	7.5	7.5	7.7	7.6
Alkalinity, mg/l $CaCO_3$	427	404	160	-	-	-	-	-	-	221	169
Chloride, mg/l	312	352	368	-	-	-	-	-	-	-	397
Total Solids, mg/l	1426	-	-	-	-	-	-	-	-	-	1327
Total Volatile Solids, mg/l	556	-	-	-	-	-	-	-	-	-	268
Total P, mg/l	11.7	9.1	1.0	-	-	-	-	-	-	-	0.7
Total Soluble P, mg/l	8.5	5.8	0.4	-	-	-	-	-	-	-	0.4
Total Ortho P, mg/l	9.2	7.2	0.5	-	-	-	-	-	-	-	0.4
Soluble Ortho P, mg/l	8.2	5.7	0.3	-	-	-	-	-	-	-	0.3
Ammonia-N, mg/l	35	37	33	-	-	32	31	33	24	9	23
Kjeldahl-N, mg/l	42	38	33	-	-	34	29	33	-	11	25
Hardness, mg/l $CaCO_3$	296	-	255	-	-	-	-	295	298	255	260
Calcium, mg/l $CaCO_3$	177	-	206	-	-	-	190	193	196	155	159
Potassium, mg/l	17	18	16	-	-	17	-	14	19	12	13
Sulfide, mg/l H_2S-S	-	-	-	trace	0.7	2.2	0.6	-	-	-	-
Dissolved Oxygen, mg/l	-	-	-	-	-	-	-	-	-	-	6.0
Chlorine Residual, mg/l	-	-	-	-	-	-	-	-	-	-	2.2
Fecal Coliform per 100 ml - geometric mean											1.7
Copper, mg/l	0.08	0.07	<0.02	<0.02	-	<0.02	<0.02	-	-	<0.02	<0.02
Nickel, mg/l	<0.05	<0.05	<0.05	<0.05	-	<0.05	<0.05	-	-	<0.05	<0.05
Lead, mg/l	<0.1	<0.1	<0.1	<0.1	-	<0.1	<0.1	-	-	<0.1	<0.1
Zinc, mg/l	0.27	0.18	<0.05	<0.05	-	<0.05	<0.05	-	-	<0.05	<0.05
Cadmium, mg/l	<0.02	<0.02	<0.02	<0.02	-	<0.02	<0.02	-	-	<0.02	<0.02
Chromium, mg/l	0.12	0.08	<0.05	0.08	-	0.06	<0.05	-	-	<0.05	<0.05
Iron, mg/l	1.43	2.3	0.81	0.45	-	0.63	0.58	-	-	0.32	0.49

[e] Approximately 10% of the BOD entering the plant was removed by contacting with the activated carbon.

[f] For those days that the ion exchange system was in service ammonia removals were substantial and averaged approximately 73%.

Because of the diameter and length of the intercepting sewer, a considerable quantity of raw sewage could be stored upstream of the plant without adversely affecting sewer users. This storage capacity was used on 55 separate occasions for a total of 388 hours when the influent gate was closed. Although this operating advantage may have been used more than absolutely necessary the above values are at least a semi quantitative evaluation of overall plant reliability.

2. Mode I

The analytical data generated during this eight week period are summarized in TABLE 30 and Figure 20. The treatment goals for BOD, suspended solids, and total phosphorus were met and removal efficiencies averaged 92%, 98%, and 95% respectively. The removal of total BOD through the carbon columns was quite low. This was not unexpected because the columns were operated in the upflow mode and thus discharged suspended solids to the filters. However, the data indicate an average removal of only 6 mg/l of BOD through the carbon columns and second stage filters. This apparent low removal is due to at least two factors. The concentration driving force for adsorption was low and, perhaps more important in terms of average values, a plant upset occurred which caused the BOD concentration of the carbon column effluent to increase substantially. On June 29, 1975 at approximately 1800 hr the pH meter servicing the pH controller for clarifier effluent was inadvertently left in the standby position after the pH and reference electrodes were cleaned. This error was not discovered until June 30 at 1100 hr. During the interim the pH downstream of the clarifier was approximately 10.5. The BOD of the GCC sample for June 30 was 44 mg/l. This extremely high value was no doubt caused by desorption of organics from the activated carbon under the high pH conditions.

No significant NH_3-N removals were accomplished because the zeolite columns were in service for only 22 days and an efficient method of regenerating the zeolite had not yet been devised. The data presented in Figure 21 illustrate the performance of the zeolite columns during period number 1 as described in TABLE 26. Both columns 4 and 5 were freshly regenerated and operated in series mode for approximately 52 hours and samples were collected at approximately 2 hr intervals.

3. Mode II

The Mode II performance data are summarized in TABLE 31 and Figure 22. The BOD, suspended solids and total phosphorus removals averaged 93%, 98% and 97% respectively. Plant performance did not suffer when the first stage dual media filters were taken out of service however operational problems were compounded. For this reason the Mode II configuration was used for only 34 days.

TABLE 30. PERFORMANCE DATA SUMMARY - MODE I

	Average Concentrations										
Parameter	Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	151	-	21	16	-	14	10	-	-	-	12
BOD - soluble, mg/l	39	-	18	-	-	10	-	-	-	-	8
COD - total, mg/l	393	-	67	57	-	42	44	-	-	-	36
COD - soluble, mg/l	111	-	47	44	-	26	-	-	-	-	25
TOC - total, mg/l	78	-	19	16	-	10	8	-	-	-	10
TOC - soluble, mg/l	31	-	15	14	-	8	-	-	-	-	9
Suspended Solids, mg/l	193	-	9	5	-	7	5	-	-	-	4
Turbidity, NTU	-	-	6	4	-	6	5	-	-	-	5
Settleable Solids, ml/l	6	-	-	-	-	-	-	-	-	-	0
Temperature, °F	55.1	-	-	-	-	-	-	-	-	-	62.0
pH	7.7	-	7.5	7.6	-	7.5	7.4	-	-	-	7.7
Alkalinity, mg/l $CaCO_3$	397	-	79	-	-	-	-	-	-	-	75
Chloride, mg/l	267	-	314	-	-	-	-	-	-	-	351
Total Solids, mg/l	1719	-	-	-	-	-	-	-	-	-	943
Total Volatile Solids, mg/l	409	-	-	-	-	-	-	-	-	-	127
Total P, mg/l	11.8	-	0.6	-	-	-	-	-	-	-	0.6
Total Soluble P, mg/l	9.1		0.4	-	-	-	-	-	-	-	0.4
Total Ortho P, mg/l	9.2	-	0.3	-	-	-	-	-	-	-	0.3
Soluble Ortho P, mg/l	9.0	-	0.2	-	-	-	-	-	-	-	0.2
Ammonia-N, mg/l	28	-	24	-	-	-	-	-	-	-	22
Kjeldahl-N, mg/l	34	-	26	-	-	-	-	-	-	-	23
Hardness, mg/l $CaCO_3$	295	-	-	-	-	-	-	-	-	-	107
Calcium, mg/l $CaCO_3$	-	-	-	-	-	-	-	-	-	-	-
Potassium, mg/l	-	-	-	-	-	-	-	-	-	-	-
Sulfide, mg/l H_2S-S	-	-	-	-	-	-	-	-	-	-	-
Dissolved Oxygen, mg/l	-	-	-	-	-	-	-	-	-	-	6.3
Chlorine Residual, mg/l	-	-	-	-	-	-	-	-	-	-	1.2
Fecal Coliform per 100/ml - Geometric Mean			-	-	-	-	-	-	-	-	1.5

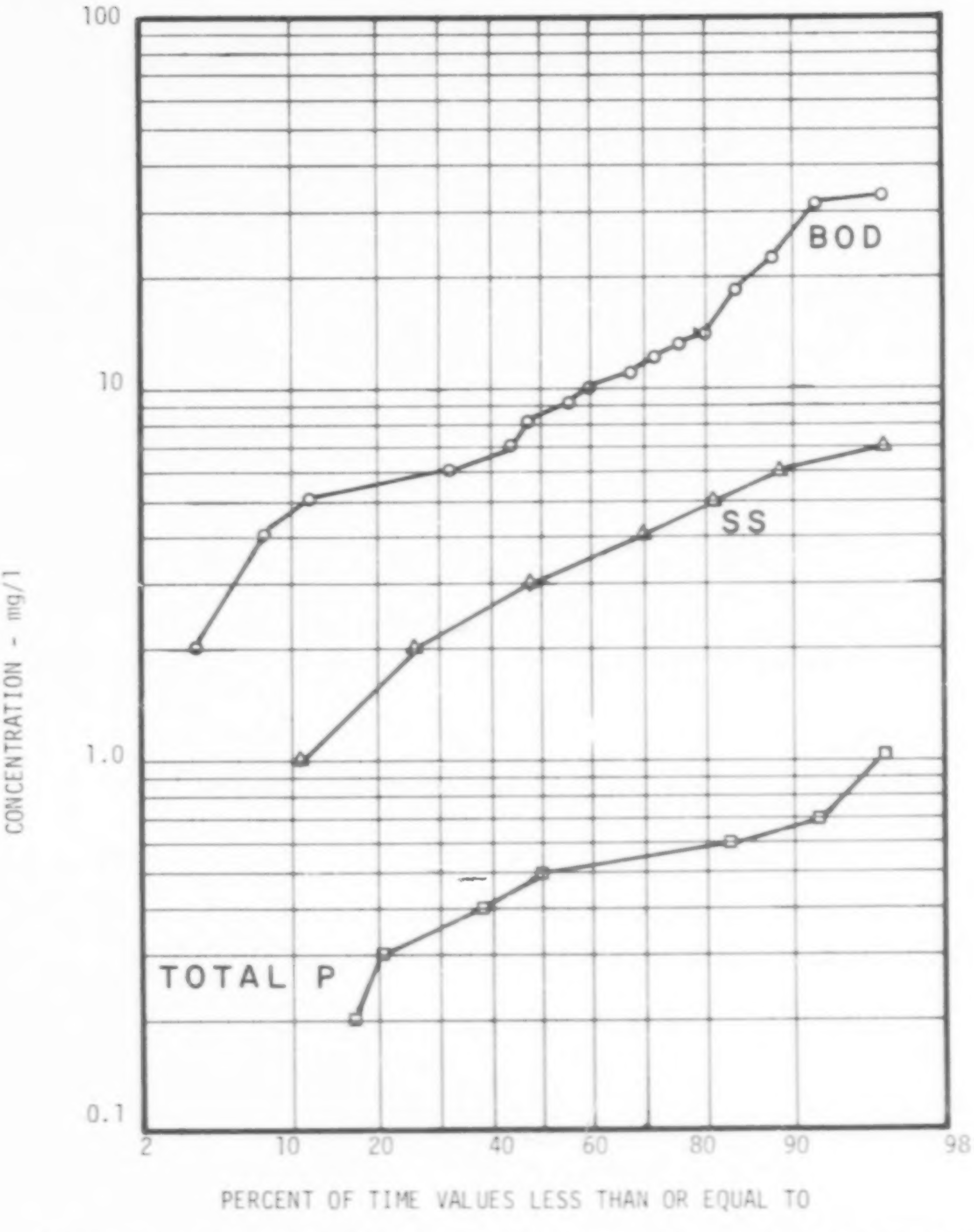


Figure 20. Frequency distributions for effluent parameters during Mode I.

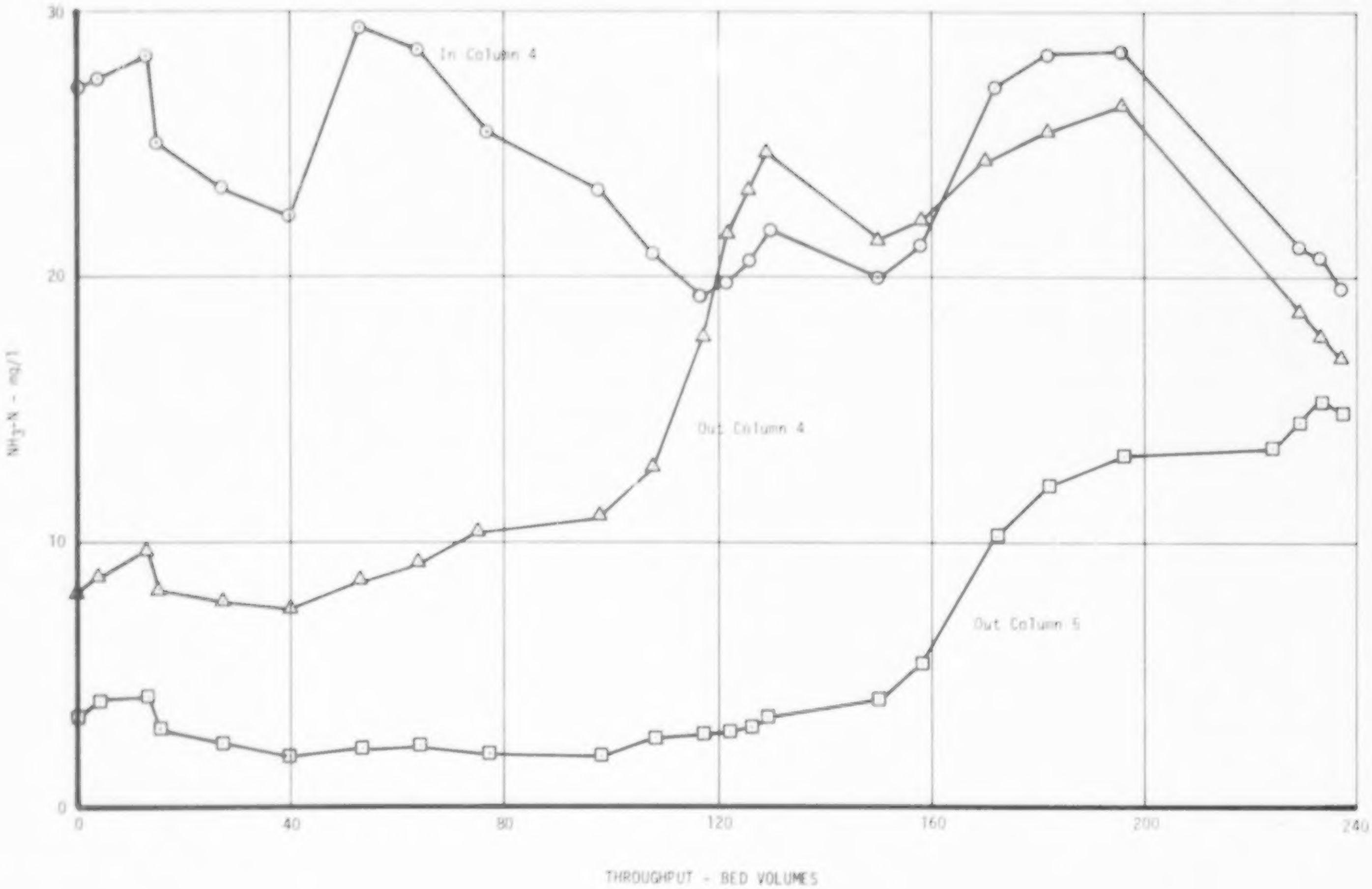


Figure 21. Performance characteristics of zeolite columns when regenerated at pH 11.

TABLE 31. PERFORMANCE DATA SUMMARY - MODE II

Parameter	Average Concentrations										
	Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	194	-	25	-	-	22	12	-	-	-	13
BOD - soluble, mg/l	53	-	20	-	-	11	-	-	-	-	11
COD - total, mg/l	491	-	70	-	-	42	32	-	-	-	37
COD - soluble, mg/l	126	-	51	-	-	22	-	-	-	-	28
TOC - total, mg/l	98	-	17	-	-	8	9	-	-	-	12
TOC - soluble, mg/l	37	-	16	-	-	8	-	-	-	-	10
Suspended Solids, mg/l	240	-	8	-	-	9	6	-	-	-	5
Turbidity, NTU	-	-	4	-	-	7	5	-	-	-	6
Settleable Solids, ml/l	8	-	-	-	-	-	-	-	-	-	0
Temperature, °F	60.4	-	-	-	-	-	-	-	-	-	65.2
pH	7.7	-	7.6	-	-	7.5	7.5	-	-	-	7.7
Alkalinity, mg/l $CaCO_3$	413	-	79	-	-	-	-	-	-	-	81
Chloride, mg/l	296	-	305	-	-	-	-	-	-	-	316
Total Solids, mg/l	1285	-	-	-	-	-	-	-	-	-	1155
Total Volatile Solids, mg/l	373	-	-	-	-	-	-	-	-	-	128
Total P, mg/l	12.2	-	0.4	-	-	-	-	-	-	-	0.4
Total Soluble P, mg/l	8.8	-	0.2	-	-	-	-	-	-	-	0.2
Total Ortho P, mg/l	9.5	-	0.2	-	-	-	-	-	-	-	0.3
Soluble Ortho P, mg/l	8.5	-	0.1	-	-	-	-	-	-	-	0.2
Ammonia-N, mg/l	32	-	33	-	-	-	-	-	-	-	26
Kjeldahl-N, mg/l	40	-	30	-	-	-	-	-	-	-	26
Hardness, mg/l $CaCO_3$	311	-	-	-	-	-	-	-	-	-	191
Calcium, mg/l $CaCO_3$	-	-	-	-	-	-	-	-	-	-	-
Potassium, mg/l	-	-	-	-	-	-	-	-	-	-	-
Sulfide, mg/l H_2S-S	-	-	-	-	-	-	-	-	-	-	-
Dissolved Oxygen, mg/l	-	-	-	-	-	-	-	-	-	-	6.8
Chlorine Residual, mg/l	-	-	-	-	-	-	-	-	-	-	0.8
Fecal Coliform per 100/ml -	Geometric Mean		-	-	-	-	-	-	-	-	2.6

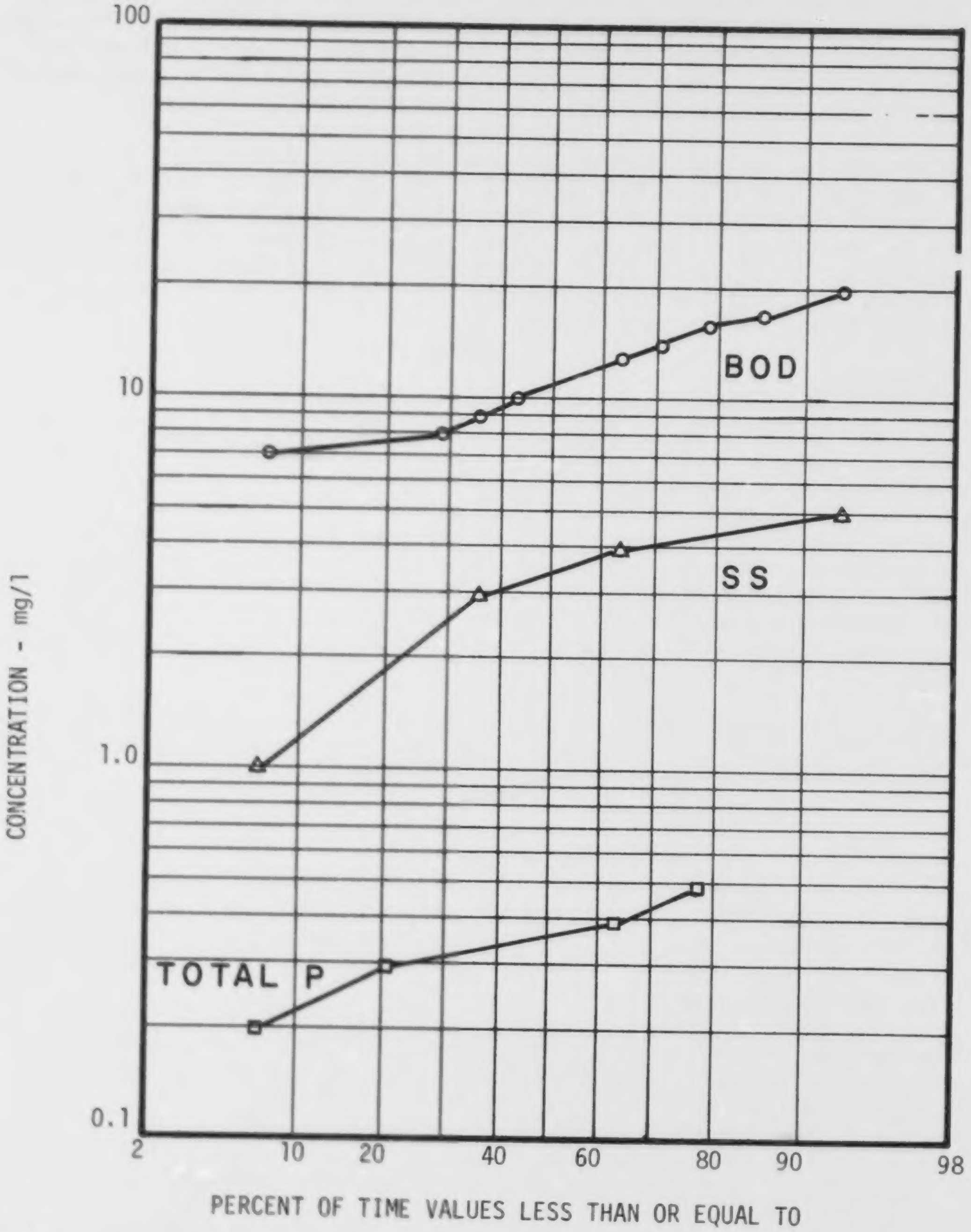


Figure 22. Frequency distributions for effluent parameters during Mode II.

The data in TABLE 31 indicate significant removals of COD and TOC through the carbon columns, however, essentially no total BOD removal. Most of the observed removal of total BOD occurred across the second stage filters. The data present a second anomaly; the BOD, COD, and TOC values in Eff were all greater than in DMF_2. The sampling system was examined several times during this period and no malfunctions or cross connections were found.

Although the zeolite columns were not in service during this reporting period approximately 20% of the influent NH_3-N was removed. The removal occurred downstream of the clarifier and was most likely related to the production of biomass in the carbon columns and filters. It is possible that some nitrification took place in filtered water storage tank No. 2 (Figure 2); however, the NO_3-N concentration of the plant effluent was determined on an irregular basis and no significant values were found.

4. Mode III

The concentration of oxidizable organics (BOD, COD, TOC) in the plant effluent was somewhat higher during this reporting period than in the two previous periods, however, the raw waste strength was substantially higher also - see TABLE 32 and Figure 23. The removal efficiencies did not change substantially compared to modes I and II and were 91%, 98%, and 96% for BOD, suspended solids and total phosphorus respectively. Substantial removals of COD and TOC were accomplished across the carbon columns but, as in mode II, no BOD removal was observed. The observed BOD removal took place in the second stage filters. The effluent values for BOD, COD, and TOC were again all higher than the corresponding values for the DMF_2 sample.

The zeolite columns were in service for 15 days during the 78 day reporting period. For those 15 days of operation the ammonia removal efficiency at the plant averaged 71% - 37 mg/l in and 10.5 mg/l out.

5. Mode IV

Plant operation during mode IV differed from previous periods in two respects; the clarifier was operated at a pH of 9.5 for most of the period and the carbon columns were operated in the downflow mode. These changes did not substantially affect plant removal efficiencies as illustrated by the data in TABLE 33 and Figure 24.

Several modifications were made to the clarifier to improve mixing in the flocculation zone, however, no performance improvements were observed and, as a result, the modifications were eliminated.

Several plant upsets occurred because of modifications in the sludge disposal procedures. During the latter part of the summer of 1975 and during the summer of 1976 the sludge generated at the Rosemount plant was disposed of by shallow sub-surface injection over an area of several acres adjacent to the plant. The sludge generated in the clarifier and zeolite regeneration area was pumped from the sludge storage tank once or twice per week as required.

TABLE 32. PERFORMANCE DATA SUMMARY - MODE III

Parameter	Average Concentration										
	Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	222	-	34	22	-	21	14	-	-	-	19
BOD - soluble, mg/l	60	-	20	-	-	14	-	-	-	-	18
COD - total, mg/l	603	-	79	67	-	50	44	-	-	-	47
COD - soluble, mg/l	146	-	51	49	-	30	-	-	-	-	38
TOC - total, mg/l	105	-	20	18	-	11	10	-	-	-	14
TOC - soluble, mg/l	41	-	16	15	-	10	-	-	-	-	13
Suspended Solids, mg/l	325	-	16	8	-	9	7	-	-	-	5
Turbidity, NTU	-	-	9	9	-	14	10	-	-	-	10
Settleable Solids, ml/l	9	-	-	-	-	-	-	-	-	-	0
Temperature, °F	60.6	-	-	-	-	-	-	-	-	-	63.5
pH	7.6	-	7.6	7.5	-	7.5	7.6	-	-	-	7.7
Alkalinity, mg/l $CaCO_3$	438	-	109	-	-	-	-	-	-	-	116
Chloride, mg/l	317	-	380	-	-	-	-	-	-	-	412
Total Solids, mg/l	1293	-	-	-	-	-	-	-	-	-	1150
Total Volatile Solids, mg/l	481	-	-	-	-	-	-	-	-	-	170
Total P, mg/l	13.0	-	0.4	-	-	-	-	-	-	-	0.5
Total Soluble P, mg/l	9.6	-	0.2	-	-	-	-	-	-	-	0.2
Total Ortho P, mg/l	10.7	-	0.1	-	-	-	-	-	-	-	0.2
Soluble Ortho P, mg/l	9.6	-	0.1	-	-	-	-	-	-	-	0.1
Ammonia-N, mg/l	36	-	43	-	-	-	-	-	-	-	25
Kjeldahl-N, mg/l	48	-	34	-	-	-	-	-	-	-	27
Hardness, mg/l $CaCO_3$	316	-	-	-	-	-	-	-	-	-	223
Calcium, mg/l $CaCO_3$	-	-	-	-	-	-	-	-	-	-	-
Potassium, mg/l	-	-	-	-	-	-	-	-	-	-	-
Sulfide, mg/l H_2S-S	-	-	-	0.1	-	8.7	-	-	-	-	-
Dissolved Oxygen, mg/l	-	-	-	-	-	-	-	-	-	-	5.6
Chlorine Residual, mg/l	-	-	-	-	-	-	-	-	-	-	1.3
Fecal Coliform per 100/ml - Geometric Mean			-	-	-	-	-	-	-	-	2.8

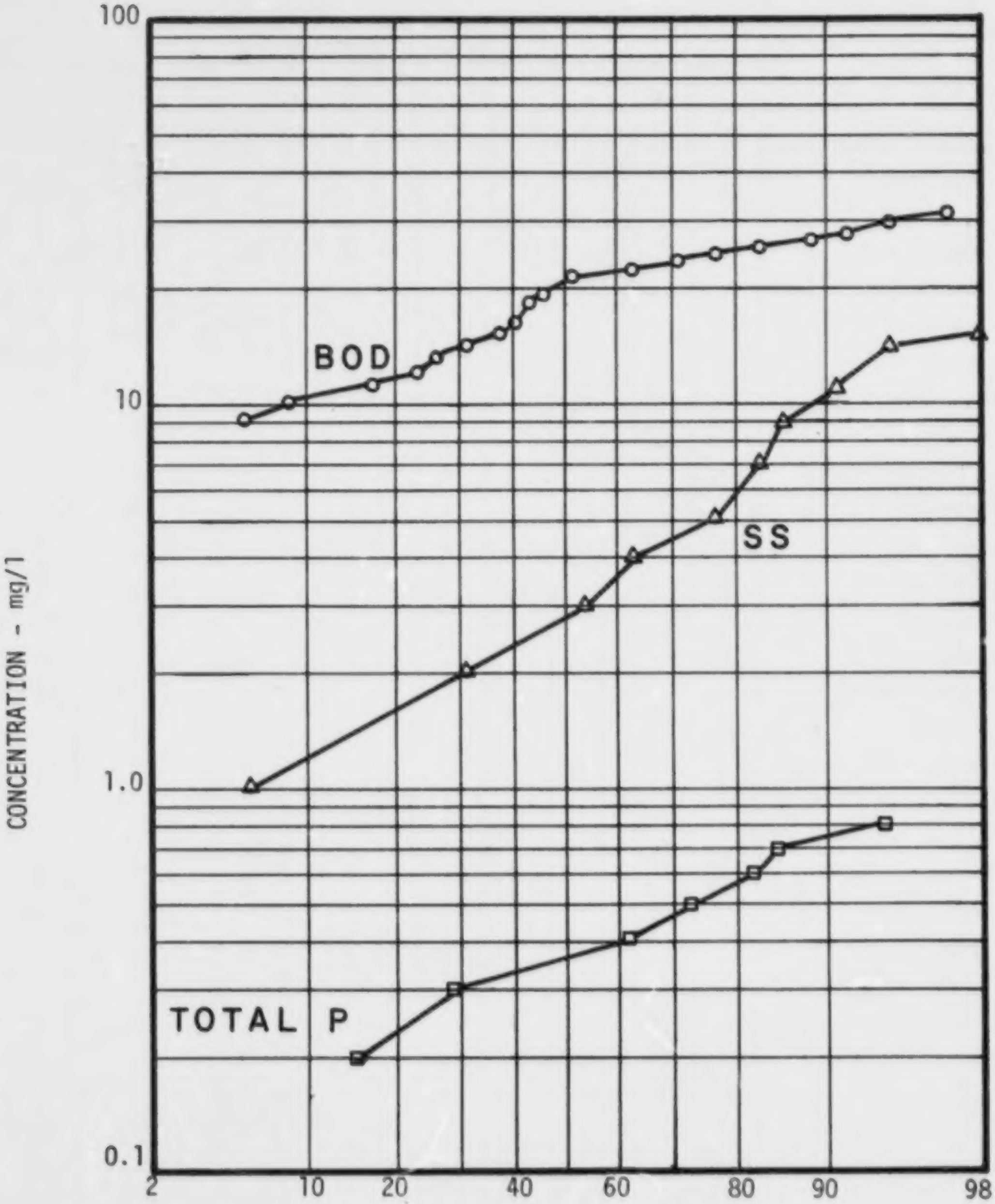


Figure 23. Frequency distributions for effluent parameters during Mode III.

TABLE 33. PERFORMANCE DATA SUMMARY - MODE IV

Parameter	Average Concentration										
	Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	221	175	46	34	-	21	15	-	-	-	15
BOD - soluble, mg/l	61	47	25	25	-	12	-	-	-	-	14
COD - total, mg/l	521	454	95	77	-	38	34	-	-	-	40
COD - soluble, mg/l	145	117	64	60	-	27	-	-	-	-	33
TOC - total, mg/l	104	89	26	21	-	10	9	-	-	-	10
TOC - soluble, mg/l	42	31	19	17	-	8	-	-	-	-	10
Suspended Solids, mg/l	237	283	22	11	-	7	5	-	-	3	3
Turbidity, NTU	-	-	10	7	-	7	7	-	-	6	8
Settleable Solids, ml/l	7	-	-	-	-	-	-	-	-	-	0
Temperature, °F	55.8	-	-	-	-	-	-	-	-	-	59.5
pH	7.6	7.8	7.5	7.5	-	7.5	7.6	-	-	7.8	7.7
Alkalinity, mg/l $CaCO_3$	434	411	184	-	-	-	-	-	-	-	196
Chloride, mg/l	315	351	374	-	-	-	-	-	-	-	405
Total Solids, mg/l	1464	-	-	-	-	-	-	-	-	-	1384
Total Volatile Solids, mg/l	583	-	-	-	-	-	-	-	-	-	293
Total P, mg/l	12.7	10.2	1.1	-	-	-	-	-	-	-	0.8
Total Soluble P, mg/l	9.2	6.6	0.5	-	-	-	-	-	-	-	0.5
Total Ortho P, mg/l	9.9	8.0	0.6	-	-	-	-	-	-	-	0.5
Soluble Ortho P, mg/l	8.9	6.4	0.5	-	-	-	-	-	-	-	0.4
Ammonia-N, mg/l	37	37	-	-	-	32	32	-	-	10	25
Kjeldahl-N, mg/l	45	40	34	-	-	-	30	-	-	12	27
Hardness, mg/l $CaCO_3$	290	-	-	-	-	-	-	-	-	-	281
Calcium, mg/l $CaCO_3$	161	-	-	-	-	-	196	-	-	166	181
Potassium, mg/l	18	18	16	-	-	17	16	-	-	12	16
Sulfide, mg/l H_2S-S	-	-	-	trace	-	2.4	0.6	-	-	-	-
Dissolved Oxygen, mg/l	-	-	-	-	-	-	-	-	-	-	5.8
Chlorine Residual, mg/l	-	-	-	-	-	-	-	-	-	-	2.4
Fecal Coliform per 100/ml -	Geometric Mean		-	-	-	-	-	-	-	-	2.0

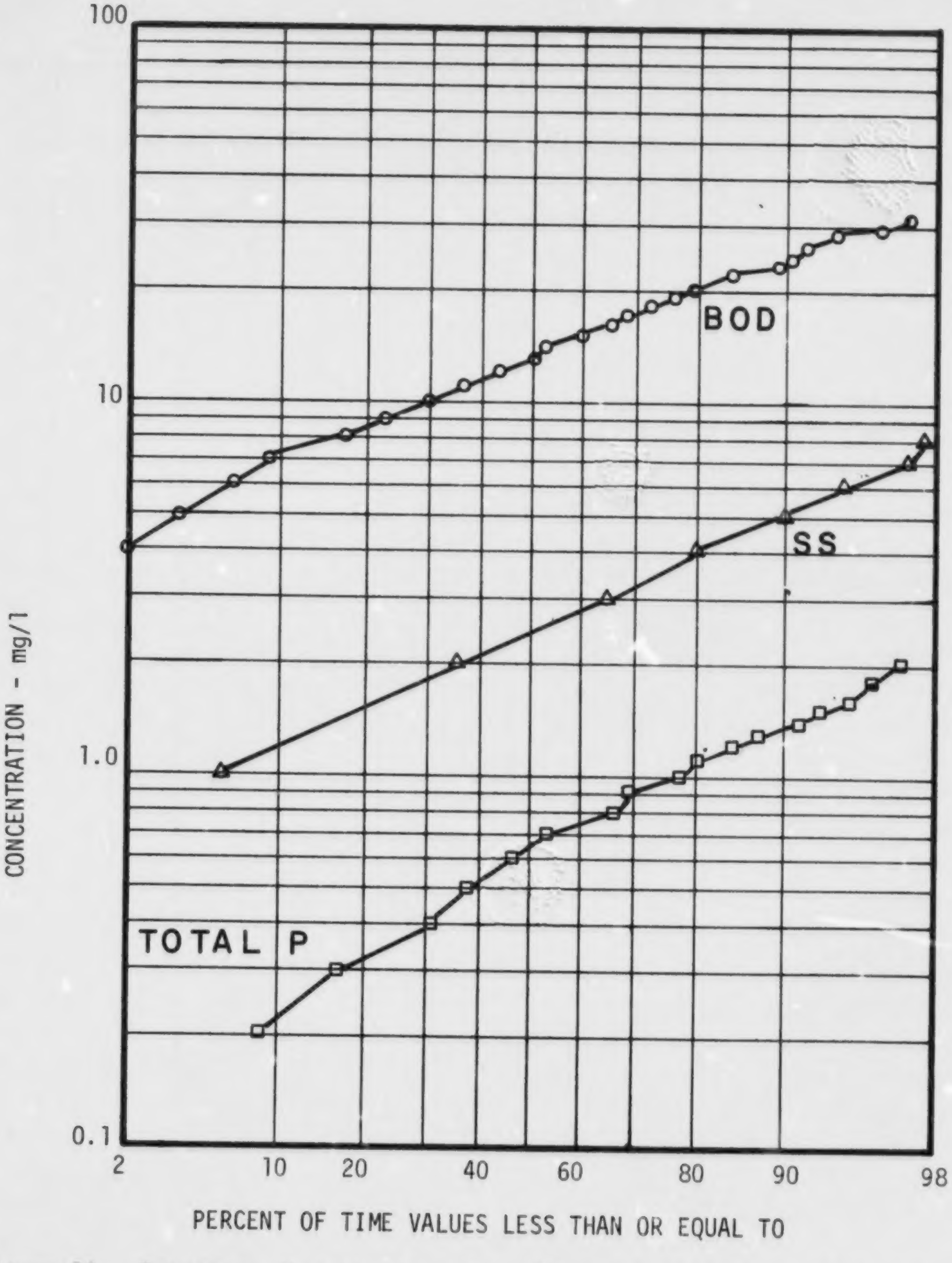


Figure 24. Frequency distributions for effluent parameters during Mode IV.

Starting August 1, 1976 the sludge was allowed to settle in the storage tank (mixer off) and the supernatant recycled to the treatment units. The decanting procedure substantially decreased the volume of the sludge removed from the plant.

Because the sludge injection equipment was needed at another plant, the last injection was accomplished August 24, 1976. After that time the sludge was hauled to another MWCC facility for ultimate disposal. Between August 24 and October 7, 32,000 gal of supernatant were returned to the plant and 40,500 gal of sludge hauled. Up to that time no difficulty was encountered because of the additional recycle flow.

On October 10 approximately 7000 gal of supernatant were recycled. An upset condition was observed in the clarifier starting at 2300 hr on October 10 and continuing through the next day. The sludge bed level in the settling zone of the clarifier rose dramatically. Approximately 8000 gal of supernatant were recycled on October 12 and the upset occurred again on the following day. Similar reactions occurred after supernatant recycle on October 24 and 25. It was observed that the pH of the stored sludge was in the range of 6 to 7 although the clarifier was operated at pH 9.5 and the sludge from the zeolite regeneration area had a pH of 11.5 to 12. Evidently the stored sludge was undergoing anaerobic digestion and the end or by-product of this reaction caused the clarifier upsets. The upsets did not occur during the period sludge was injected; however, the sludge storage tank was completely pumped out during each injection. During the period that sludge was hauled the sludge storage tank was never completely emptied and it appears that sufficient seed organisms remained to ensure a continuous digestion process. For this reason the attempt to thicken the Rosemount sludge prior to disposal was terminated on October 25, 1976.

During this reporting period no apparent increase in BOD or TOC between the GCC and Eff samples was observed.

Although the zeolite columns were in service for a total of 123 days or approximately 30% of the time the removal efficiency for NH_3-N averaged only 32%. For the days the columns were in service the removals averaged approximately 73%. During that period however several regeneration procedures were evaluated.

6. Mode V

The performance data obtained during mode V, TABLE 34 and Figure 25, demonstrate that the removal of the second stage filters from the process configuration had little effect on the overall plant performance. The performance goals were met in terms of suspended solids. BOD, and total phosphorus removals. The zeolite columns were in service a total of 62 days or approximately 43% of the time. The zeolite regeneration system was kept out of service during January and February 1977 because of the energy crisis in the State of Minnesota. During that two-month period the temperature in the plant was maintained at 55°F and all nonessential fuel consumption was eliminated. For those days the zeolite columns were in service the ammonia removals averaged 79%.

TABLE 34. PERFORMANCE DATA SUMMARY - MODE V

Parameter	Average Concentration										
	Raw	Clar in	Clar	DMF1	GCC1	GCC	DMF2	In AEC	AEC1	AEC	Eff
BOD - total, mg/l	204	175	40	32	18	13	-	-	-	-	15
BOD - soluble, mg/l	65	49	28	27	14	9	-	-	-	-	14
COD - total, mg/l	459	409	100	80	48	31	-	38	-	-	38
COD - soluble, mg/l	151	118	71	65	40	24	-	-	-	-	33
TOC - total, mg/l	107	88	28	23	13	10	-	10	-	9	12
TOC - soluble, mg/l	47	36	21	19	12	8	-	-	-	-	11
Suspended Solids, mg/l	214	199	20	9	5	5	-	2	1	2	2
Turbidity, NTU	-	-	9	5	4	4	-	6	3	4	6
Settleable Solids, ml/l	7	-	-	-	-	-	-	-	-	-	0
Temperature, °F	50.9	-	-	-	-	-	-	-	-	-	56.8
pH	7.6	7.9	7.3	7.3	7.5	7.4	-	7.5	7.5	7.7	7.5
Alkalinity, mg/l $CaCO_3$	417	391	170	-	-	-	-	-	-	-	184
Chloride, mg/l	318	353	379	-	-	-	-	-	-	-	401
Total Solids, mg/l	1405	-	-	-	-	-	-	-	-	-	1352
Total Volatile Solids, mg/l	547	-	-	-	-	-	-	-	-	-	276
Total P, mg/l	8.1	7.2	1.0	-	-	0.4	-	-	-	0.4	0.5
Total Soluble P, mg/l	5.7	4.5	0.3	-	-	-	-	-	-	-	0.3
Total Ortho P, mg/l	6.4	5.4	0.4	-	-	-	-	-	-	-	0.4
Soluble Ortho P, mg/l	5.4	4.1	0.2	-	-	-	-	-	-	-	0.2
Ammonia-N, mg/l	32	-	-	-	-	32	-	33	24	7	15
Kjeldahl-N, mg/l	38	35	33	-	-	33	-	33	-	9	17
Hardness, mg/l $CaCO_3$	279	-		-	-	260	-	295	298	255	267
Calcium, mg/l $CaCO_3$	187	-		-	-	184	-	193	196	147	164
Potassium, mg/l	-	-		-	-	-	-	14	19	11	-
Sulfide, mg/l H_2S-S	-	-		trace	0.7	1.4	-	-	-	-	-
Dissolved Oxygen, mg/l	-	-		-	-	-	-	-	-	-	6.6
Chlorine Residual, mg/l	-	-		-	-	-	-	-	-	-	2.8
Fecal Coliform per 100/ml -	Geometric Mea			-	-	-	-	-	-	-	1.4

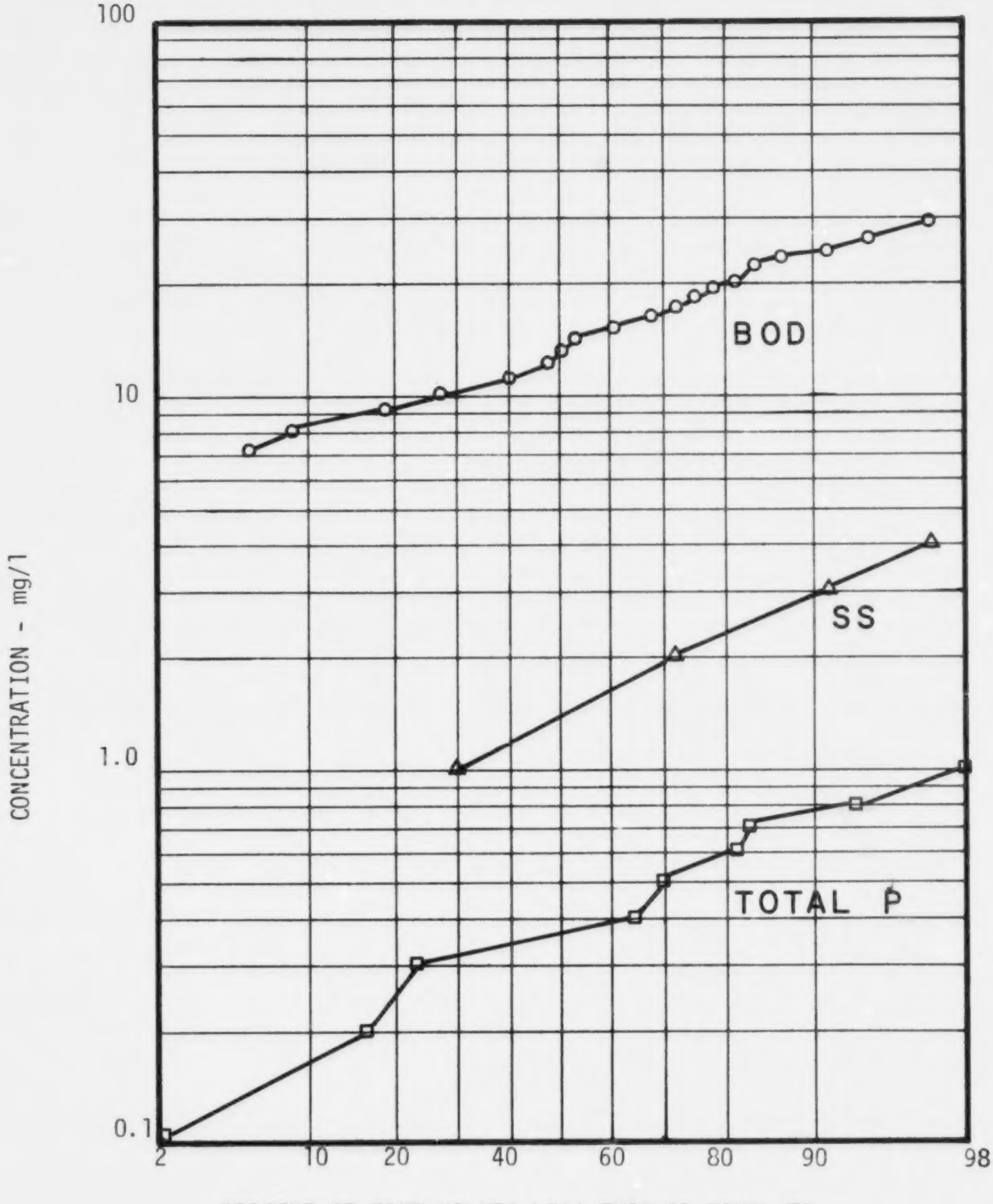


Figure 25. Frequency distributions for effluent parameters during Mode V.

The average concentration of organics in the effluent (measured by BOD, COD, and TOC) was again greater than the average concentration discharged from the carbon columns. Because zero is the lower unit and the upper range is generally not constrained the concentrations of pollutants in wastewater are usually not normally distributed. The logarithms of the concentration values are however often normally distributed. (9)(10)(11) The plant effluent data presented in Figures 19, 20, 22, 23, 24, and 25 indicate that the log normal distribution is an appropriate description for the concentration data.

The student's t test (12) was used to compare the means of the two sample populations. The hypothesis $u_x = u_y$ was tested against the alternative $u_x \neq u_y$ where u_y is the average value for an effluent parameter and u_x is the average value of the parameter for the DMF_2 or GCC samples. The type one error (the probability that $u_x = u_y$ but rejected) was fixed at 5%.

The tests were conducted for the data collected by mode assuming both normal and lognormal distributions. Because the number of observations for the Eff sample was usually greater than for the DMF_2 or GCC sample the tests were duplicated for paired data only. The results summarized in TABLE 35 indicate that for modes III, IV, and V, which account for 88% of the demonstration period, the organic content of the plant flow increased significantly between the last treatment process and the plant effluent sampling point.

On February 22 and 23, 1977 grab samples were collected to determine if the composite samples were affected by transport from the point of sampling to the sample compositing apparatus. The following samples were collected.

[1] effluent - collected at discharge of chlorinator feed pump, see TABLE 4

[2] effluent - collected in sampling room, see Figure 2

[3] GCC - sampled from bottom tap on secondary carbon column

[4] GCC - sampled in sampling room, see Figure 2 and TABLE 4

Sample [2] was transported through approximately 60 ft of 1/2 in polyethylene tubing and 70 ft of 3/4 in PVC pipe. Sample [4] was transported through approximately 80 ft of 3/4 in PVC pipe. Grab samples were collected on an hourly basis and TOC analyses conducted. The results of this test, presented in Figure 26, indicate that sample transport had no significant effect on the measured organic content of the samples.

Further linear regression analysis of the data presented in Figure 26 indicate the the TOC values for the GCC and Eff samples are related as follows:

For samples [2] and [4] Eff = 0.55 + 1.03 GCC
For samples [1] and [3] Eff = 0.71 + 0.99 GCC

TABLE 35. COMPARISON OF POPULATION MEANS - Eff vs DMF_2, GCC

Hypothesis: $u_x = u_y$; Type 1 error = 5%

u_y = average value of parameter for Eff sample

u_x = average value of parameter for DMF_2 sample except for mode V where GCC sample is used

		Status of Hypothesis				
		Mode				
Distribution Assumed	Parameter	I	II	III	IV	V
Lognormal	BOD	accept	accept	reject	accept	reject
	COD	accept	accept	accept	reject	reject
	TOC	no data	accept	reject	accept	reject
Lognormal - Pairs	BOD	accept	accept	reject	reject	reject
	COD	accept	reject	reject	reject	reject
	TOC	no data	reject	reject	reject	reject
Normal	BOD	accept	accept	reject	accept	reject
	COD	accept	accept	accept	reject	reject
	TOC	no data	accept	reject	accept	reject
Normal - Pairs	BOD	accept	accept	reject	accept	reject
	COD	accept	accept	accept	reject	reject
	TOC	no data	accept	reject	accept	reject

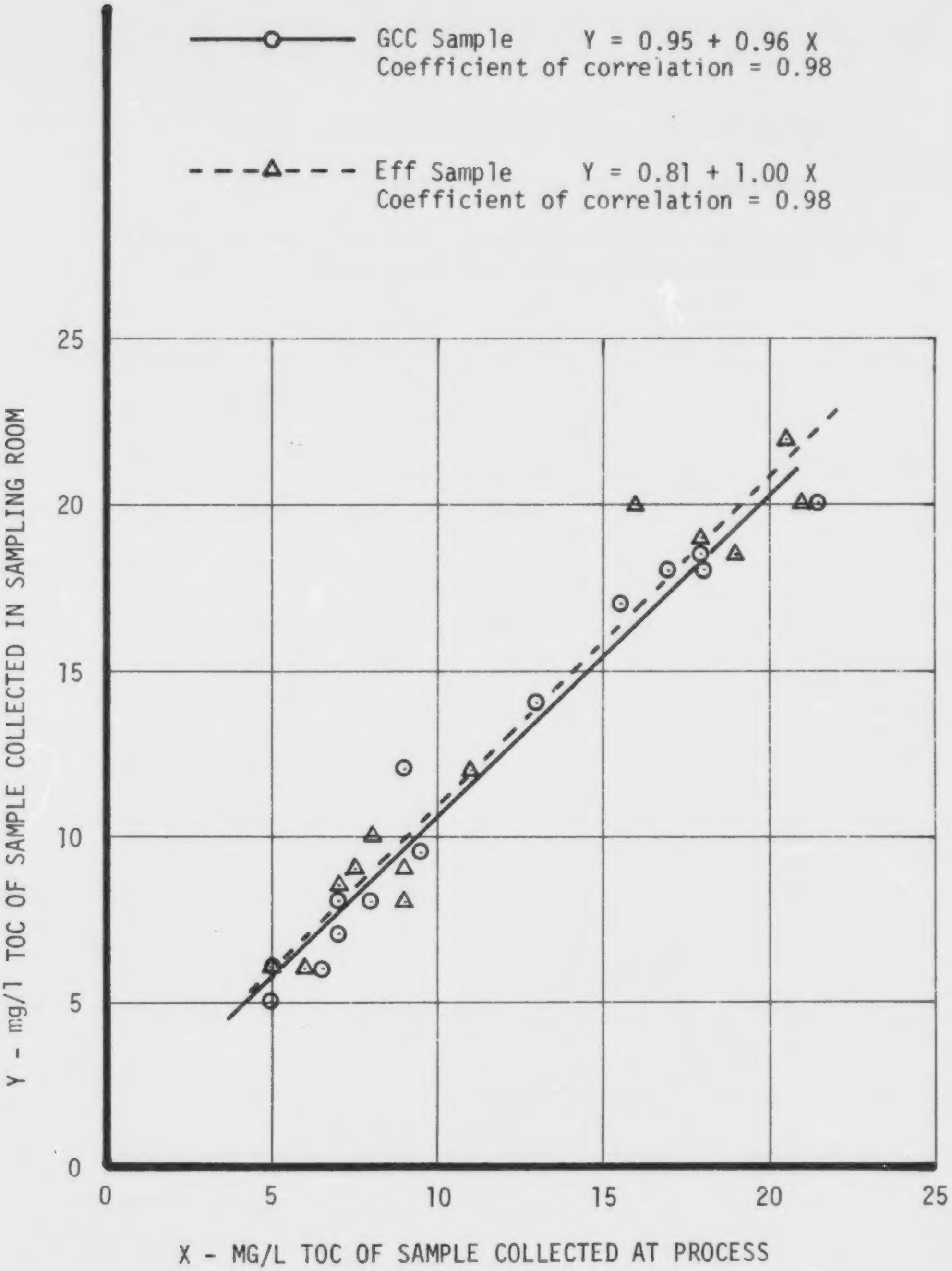


Figure 26. Effect of sample transport on concentration of TOC.

These data demonstrate that the sample transport system did not have a significant influence on the measured organic content of the sample. The data, however, do demonstrate that the organic content of both the GCC and Eff samples varied in the range of 4:1 during a days time. Variations of this magnitude were observed during several sampling programs and were apparently related to the carbon column backwash procedure. Prior to March 1, 1977 when one of the carbon columns was taken out of service for backwashing only one column remained in service and the organic content of the GCC and downstream samples increased. The apparent increase in concentration of organics between the GCC and Eff samples may have been due to the variations described along with some bias in the flow measuring/sample activation systems.

XI. UNIT PROCESS PERFORMANCE

A. CLARIFICATION

1. Operating Procedures

a. Sludge Blanket Control

The clarification system was designed with no direct mechanical energy input to the flocculation zone - Figure 2. However since the number of collisions between particles is a function of the average velocity gradient, G, the volume concentration of particles, C, and the basin detention time, T, the efficiency of flocculation can be enhanced by maintaining a high concentration of solids in the flocculation zone and/or maintaining a sludge blanket in the upflow clarification zone. During the first several months of operation it was observed that if the flow passed up through an expanded sludge blanket of approximately 30 inches the solids capture was enhanced and the effluent turbidity reduced substantially. Experience during that period also indicated however that control of the sludge blanket was extremely difficult because of the flow variations experienced at the plant - Figure 13. On several occasions prior to initiation of the demonstration project the sludge blanket discharged over the weirs and plugged the first stage filters which required the entire plant to be shut down for approximately one half of a day. Because of the above experience and that of others (6)(13)(14) the clarifier was normally operated without a sludge blanket in the upflow zone.

The bottom of the skirt dividing the flocculation and settling zones - Figure 2 - is approximately 110 in. below the normal water surface elevation. A sample tap along with a hose and cable device was used to locate the sludge-water interface to a depth of 120 in. The plant operators' normal routine included locating the interface level on an hourly basis. The distance between the water surface and the interface was normally maintained at 120 in. If for some reason the sludge level started to rise some remedial action was taken which would include adjusting the mixer speed, the sludge blowdown rate or the clarifier flow rate depending on the rate of change of the sludge level.

b. Mixer Speed

The contractor's representative initially recommended that the mixer speed be adjusted according to the results of what was identified as the V_1-V_2 test. The V value was defined as the volume of sludge below the solids-liquid interface after 10 min settling in a 100 ml graduated cylinder. It was recommended that samples be taken at two locations, the bottom of the draft tube at the level of the mixer and at the top of the draft tube. The recommendations were to maintain the V_1 value at the bottom of the draft tube at ≤ 15 and to maintain the V_2 value at the top of the draft tube at $V_1-(3 \text{ to } 5)$ by varying the mixer speed. The V-V tests and speed adjustments were to be made on a semi-hourly basis. For the lime-ferric chloride combinations used during the demonstration period this method of control did not appear to affect clarifier performance but did require continuous adjustment of the mixer speed.

During the period when the lime-ferric chloride combinations were used the mixer was adjusted to yield visible movement at the water surface near the center of the flocculation zone. The speed was adjusted down when the sludge bed moved up for more than a day and when the turbidity of the clarifier effluent increased. During the two-year demonstration period the mixer speed was operated in the range of 70 to 155 RPM and averaged 97 RPM. Because of fouling with rags and other debris the mixer RPM was not always a good indication of energy input or mixing intensity. The clarifier was drawn down and partially cleaned 14 times during the two-year period. During each cleaning operation the mixer was found to be heavily fouled with rags. Although the raw wastewater was coarse screened at 1/4 in. clear openings a sufficient quantity of rags and other stringy debris passed through to the clarifier. On several such occasions it was observed that rags and other debris completely filled the mixing zone to a depth of approximately 2 ft above the top of the mixer. Similar experiences are reported in the literature (15).

After a period of approximately 10-15 days after cleaning the mixing intensity and ability to recirculate settled solids was significantly reduced based on visual observations at the water surface.

c. Sludge Wasting

During the last half of 1974 when lime-ferric chloride combinations were used in the clarifier it was determined that the sludge wasting rate could be used to control the solids content of the waste sludge. When the sludge wasting rate was decreased the solids inventory in the clarifier increased and the blowdown concentration also increased. In September 1974 blowdown concentrations as high as 20% solids were obtained. It was also observed however that as the sludge thickened in the clarifier the rate of solids recycle decreased substantially causing increased turbidity in the discharge. The thickened sludge also caused the sludge plow drive to fail on several occasions because of overload conditions.

In order to minimize problems with solids recirculation and the possibility of upsets related to increased microbial activity the sludge wasting schedule was adjusted to maintain the blowdown concentration in the range of 5 to 10% solids. During the two-year period the waste sludge averaged 7.44% solids. This was obtained by wasting an average of 4780 gpd or approximately 12,250 lbs solids per million gal raw wastewater treated. The average blowdown rate was one cycle per 1630 gal and the average blowdown time was 9 sec.

2. Chemical Conditioning Alternatives

As previously discussed a number of combinations of lime, ferric chloride and polymer were used prior to the initiation of the demonstration project. The ferric chloride and polymer combinations were eliminated from consideration because of the carryover associated with the normal diurnal flow variations and because the sludge produced did not thicken beyond 1% in the clarifier or sludge storage tank.

Jar test data indicated that the low lime process - lime to pH 10.1 plus polymer - could be used to achieve required phosphorus removals. This scheme was used several times in the clarifier for short periods (2-3 days) during the period July 1974 to May 1975. During each attempt the suspended solids and phosphorus concentrations increased substantially and the low lime process was thus eliminated from further consideration. The failure to duplicate the treatment efficiencies obtained in the jar test was most likely related to the relative mixing regimes and to the fact that sludge was not recycled in the jars. The jar tests were conducted with apparatus as manufactured by Coffman Industries, Inc. One liter samples of raw wastewater were dosed with the lime slurry, rapid mixed at 100 RPM for 5 min., slow mixed at 50 RPM for 15 min., and settled for 15 min. The slow mix speed was sufficient to keep all solids in suspension. Based on visual observations of the jars and the liquid surface in the flocculation zone of the clarifier the mixing intensity was much greater in the jars. The mean velocity gradient, G, for similar stirring devices has been reported to be approximately 70 fps/ft at room temperature. (16)(17) It is difficult to estimate the mixing intensity in the flocculation zone because there is no input of mechanical energy. In February of 1975 samples were taken at the exit of the mixing and flocculation zones. One liter aliquots were subjected to further slow mixing at 20 RPM in the jar test apparatus, $G \approx 25$ fps/ft (16). The results presented in TABLE 36 demonstrate that the mixing intensity in the flocculation zone was not sufficient to provide for efficient flocculation.

TABLE 36. EFFECT OF ADDITIONAL SLOW MIXING ON SOLIDS AND PHOSPHORUS REMOVAL

Sample	Slow mix time * min	Turbidity	Ortho P
Exit mixing zone	0	8.9	0.37
	5	4.0	0.35
	10	3.0	0.28
	20	2.5	0.28
Exit flocculation zone	0	20	0.34
	5	8.5	0.28
	10	6.7	0.24
	20	5.1	0.24

* All jars settled 20 min. prior to sampling

Jar tests indicated that floc, assumed to be $Mg(OH)_2$, formed at a pH value of approximately 11.1 to 11.3 and that good clarification could be obtained with high lime treatment. The jar tests indicated that a lime dose of approximately 510 mg/l CaO would be required to obtain a pH value of 11.1. For a 30-day period during November and December 1974 the high lime process was used. The average of approximately 700 hourly pH values was 11.4 (the average pH was somewhat higher). In order to maintain a high degree of solids capture approximately 2.7 mg/l of Nalco 677 was added in the mixing zone. The clarifier performance data for that period were as follows: suspended solids 3.9 mg/l, turbidity 4.0 NTU, BOD 44 mg/l and total ortho phosphorus 0.7 mg/l. Although the solids removal efficiency was more than adequate the removals of phosphorus and BOD were less than anticipated. The process was eliminated from further consideration because of the extremely high lime dose required - approximately 830 mg/l CaO. This was significantly higher than anticipated from the jar test data and may have been related to the solids inventory in the clarifier (14) and to incomplete dissolution of the lime in the clarifier mixing zone. As indicated in Figure 2, the lime slurry addition point was slightly below the plane of the mixing impeller. It is possible that a significant fraction of the lime slurry was not mixed with the incoming flow but rather settled into the sludge zone directly.

Menar and Jenkins (18) recommended that ferric chloride or alum be used to coagulate precipitated calcium phosphate when the pH is less than 10.0. Stamberg et al (19) found that total phosphorus residuals exceeded 2 mg/l at pH values less than 11.5 unless a supplemental flocculant such as ferric chloride was used. Other investigations (15) (20) have demonstrated high removal efficiencies for phosphorus, suspended solids and organics when using lime and ferric chloride combinations. Based on these results and jar tests at the Rosemount facility only lime and ferric chloride combinations were considered for use after December 1974.

3. Removal Efficiencies

a. Overall

The data presented in TABLE 29 and Figures 27 and 28 demonstrate the average performance characteristics of the chemical clarification process in terms of removing suspended solids, phosphorus and organics. During this period the average hydraulic loading to the clarifier was 0.5 gpm/sq ft and varied in the range of approximately 0.1 to 0.9. Although the characteristics of the raw wastewater varied substantially even on a weekly average basis, the characteristics of the clarifier effluent were relatively constant. This damping effect is more pronounced when daily data are considered as in TABLE 37. The removal efficiencies for suspended solids, phosphorus and organics obtained at Rosemount compare favorably with data reported for pilot facilities (15)(21).

It is interesting to note that significant fractions of what are herein termed soluble materials were also removed in the clarification process. These materials were most likely not soluble but suspensions of colloidal and supra-colloidal particles. The removal of these finely divided particles was

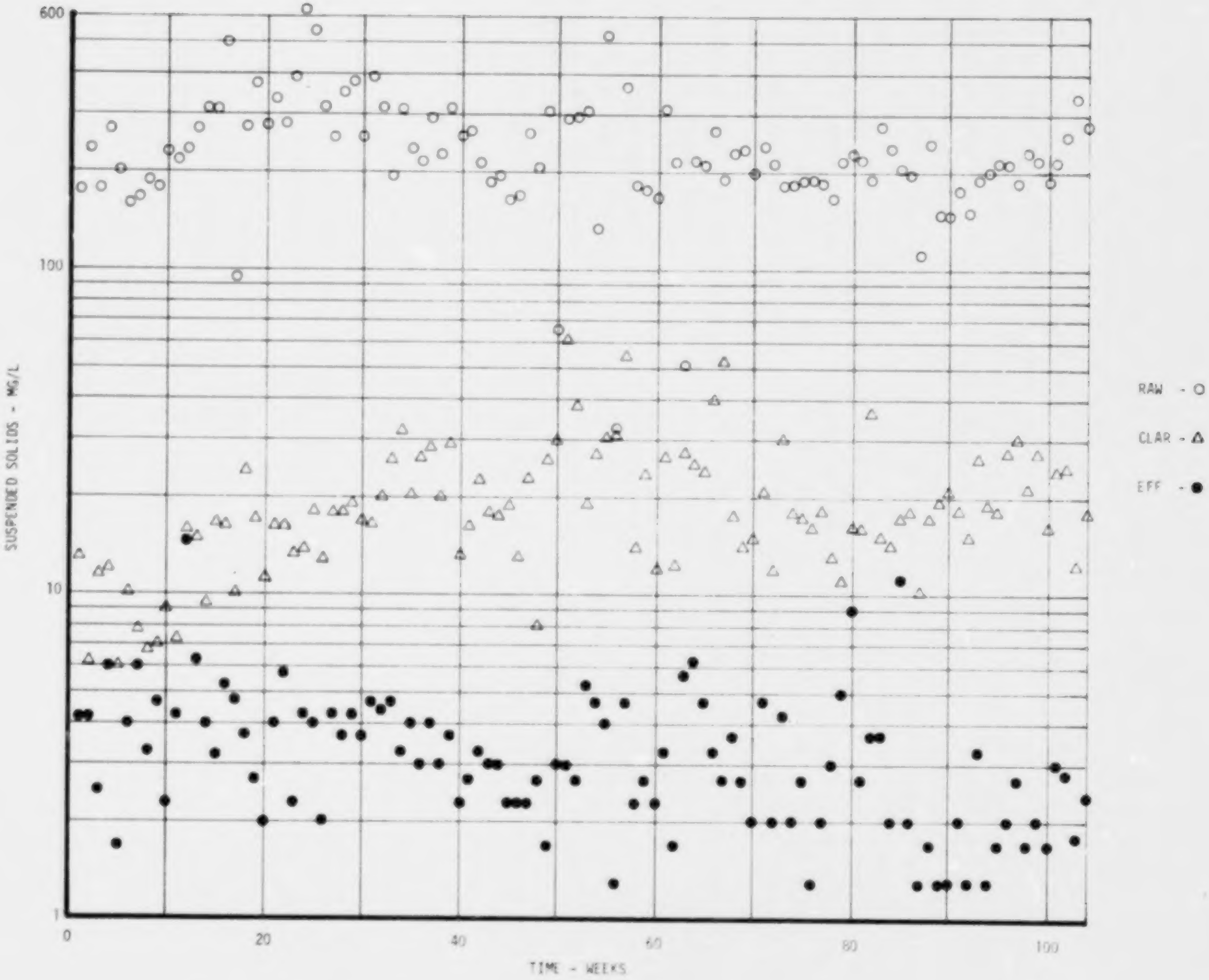


Figure 27. Weekly average suspended solids concentrations.

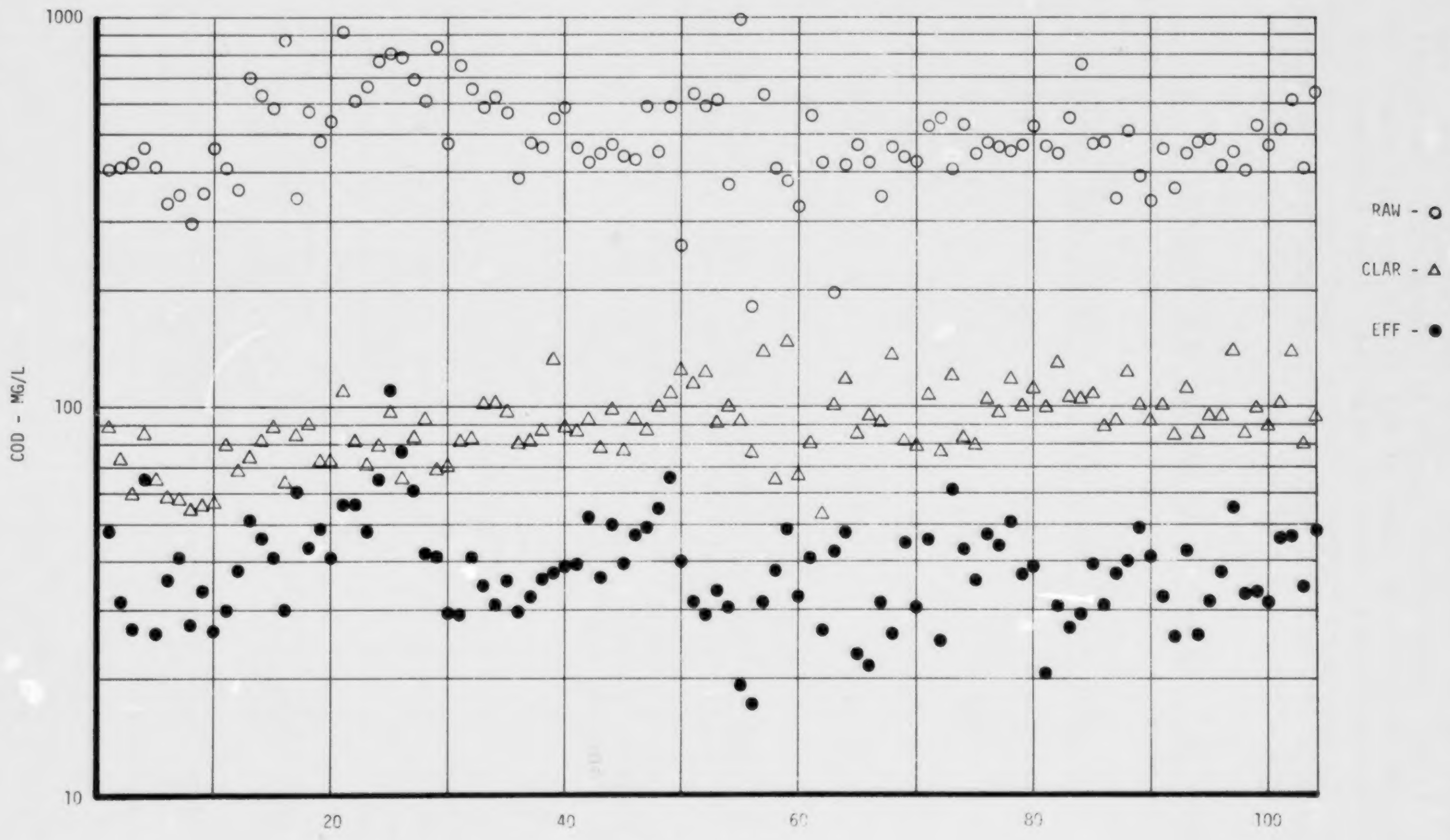


Figure 28. Weekly average COD concentrations.

most likely effected by the coagulants through the two mechanisms - adsorption and entrapment. If the proper mixing conditions could have been established in the rapid mix and flocculation zones of the clarifier the removal efficiencies of soluble constituents could no doubt have been increased and the observed variability decreased further. Several investigators have demonstrated the importance of rapid and complete initial mixing of the coagulant in terms of achieving efficient use of the applied chemicals. (22)(23).

TABLE 37. VARIABILITY OF RAW AND CLAR PARAMETERS

	Raw		Clar	
	mean	std. deviation	mean	std. deviation
SS	237	152	19	13.5
COD - total	505	234	91	26
COD - soluble	143	36	62	20
Total P	11.7	3.7	1.0	0.6

The data presented in TABLE 29 indicate that the heavy metals which were found to have concentrations greater than detection limits were efficiently removed in the clarifier. The removals were most likely accomplished through formation of the sparingly soluble metal hydroxides. Only iron was found to carry over in significant concentrations and this was most likely related to the discharge of pinpoint ferric hydroxide floc.

b. pH 10.5

Clarifier operation during the demonstration period can be divided into two operating modes based on the pH maintained in the mixing and flocculation zones. During the first six months the set point pH was 10.5. For the following 18-month period the set point pH was lowered to 9.5. As previously described the lime slurry feed pumps were controlled to maintain constant dosage. Because of the variable alkalinity of the clarifier influent - due primarily to recycle flows - constant dosage did not yield good pH control. Better pH control was obtained by manually adjusting the set point of the pump controller on an hourly basis as required. The averages of the 24 hourly pH values were calculated for each day and are presented in Figure 29. Although these are not average pH values (on a flow proportional basis) the values demonstrate to a limited degree the consistency of pH control which was one of the important process parameters.

For the first six month period the average of 4161 hourly pH values was 10.5. During this period the average hydraulic loading of the clarifier was 0.47 gpm/sq ft and the average chemical feed rates were 448 mg/l lime [CaO] and 12.4 mg/l Fe. During the first 22 weeks of the period polymer (Nalco 677)

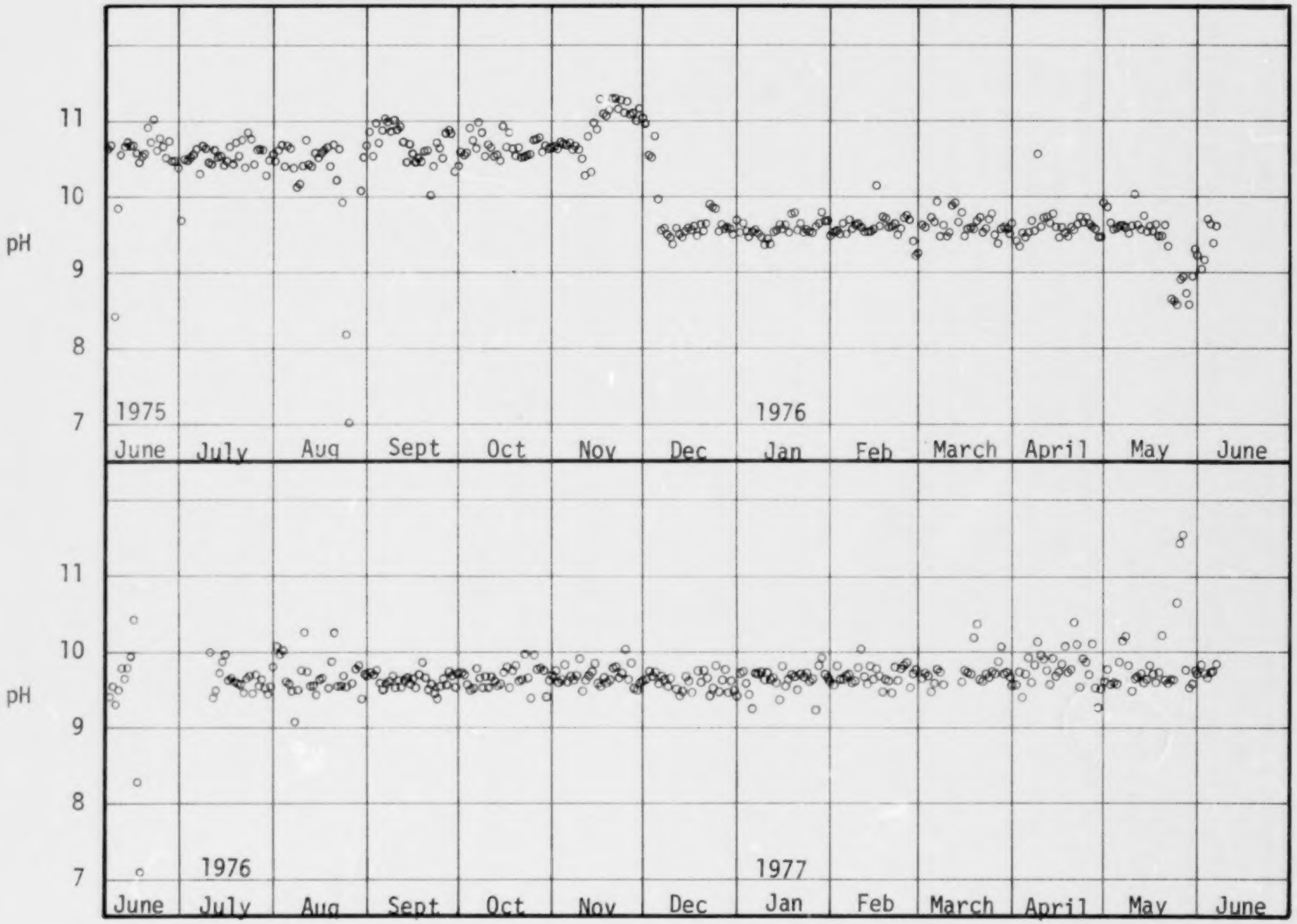


Figure 29. Average of hourly pH values for clarifier reaction zone.

was also used at a rate of approximately 2.5 mg/l. Polymer use with the lime-ferric combinations was terminated on November 16, 1975. The lime addition point was moved to the influent pipe (Figure 3) during the last week in September 1975. It was anticipated that the somewhat better mixing condition would yield more efficient lime utilization and lead to a reduction in dose. Although no significant change was observed the alternate addition point was used for the remainder of the study.

The average ferric chloride feed rates, determined on a 28 day basis varied from 2.6 to 15.1 mg/l Fe. The clarifier performance, based on the analyses of the 24 hr composite samples, appeared to be independent of the dosage over the range observed. It was observed, however, that during periods of low flow the clarifier was extremely turbid when the average ferric chloride dose was less than 7 mg/l Fe. Based on visual observations treatment efficiency increased as the flow through the unit increased. This apparent contradiction was most likely caused by increased mixing intensity in both the rapid mix and slow mix zones due to the increased input of energy.

The average performance characteristics are summarized in TABLE 38. In the State of Minnesota the normal secondary treatment standards for 30 day averages are 25 mg/l BOD and 30 mg/l suspended solids. Based on this definition secondary treatment can be nearly obtained by chemical clarification alone. The range and variability of several clarifier effluent parameters are presented in Figures 30, 31 and 32. The sludge wasting rate averaged approximxately 6500 gpd at an average total solids concentration of 5.3%. An average of 13,200 lbs solids were wasted per million gallons raw wastewater treated.

TABLE 38. CLARIFIER PERFORMANCE CHARACTERISTICS AT pH 10.5

Parameter	Average Concentrations - mg/l Raw	Clar	% Removed *
Suspended solids	270	12	94
Turbidity (NTU)	-	7	-
BOD - total	198	28	81
BOD - soluble	53	19	56
COD - total	539	73	82
COD - soluble	131	50	50
Total P	12.5	0.5	95
Total soluble P	9.3	0.2	97
Kjeldahl-N	42	30	6.5

* based on mass flow rates in raw and clarifier flows

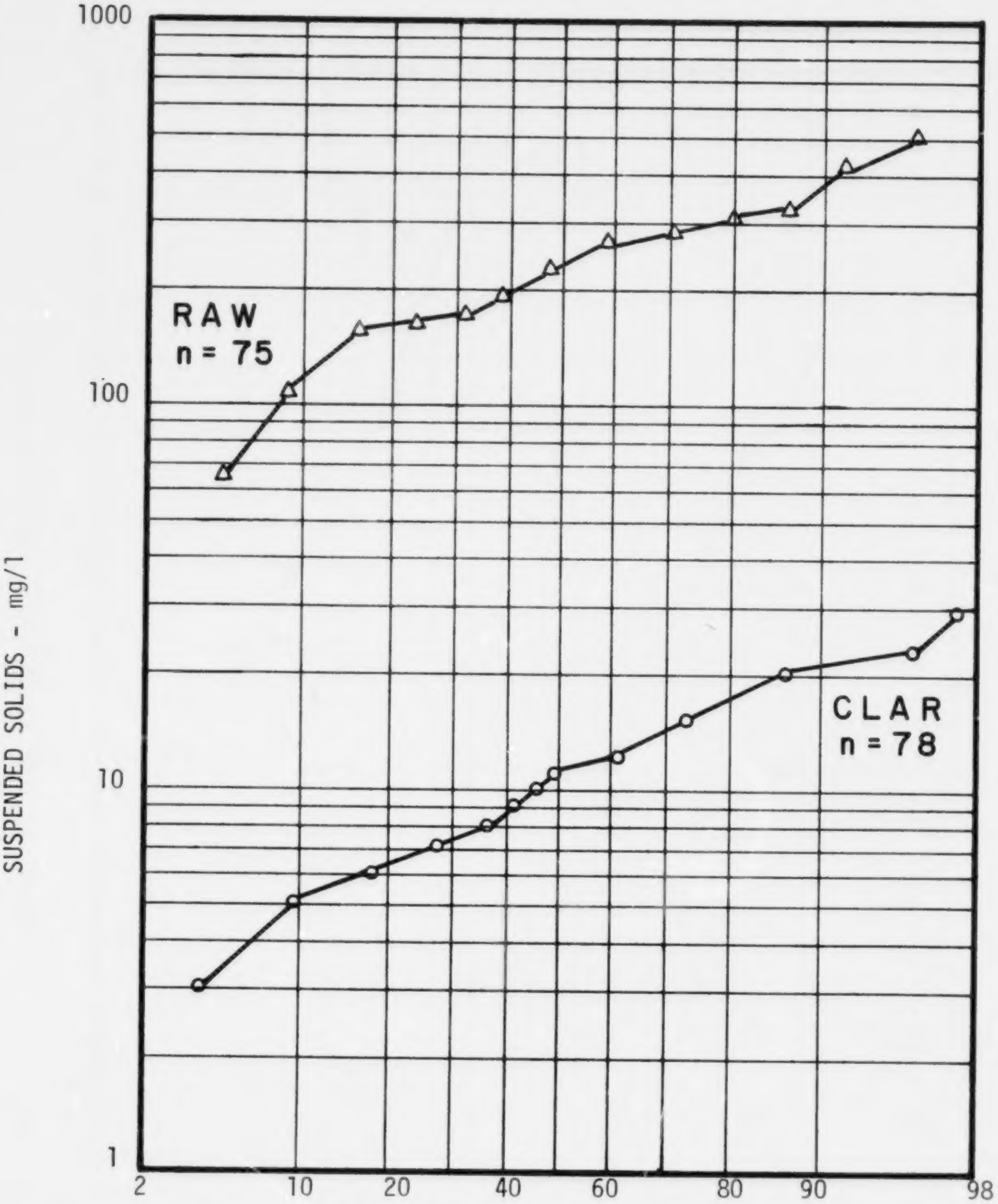


Figure 30. Frequency distributions for Suspended Solids data - 6/7/75 through 12/5/75.

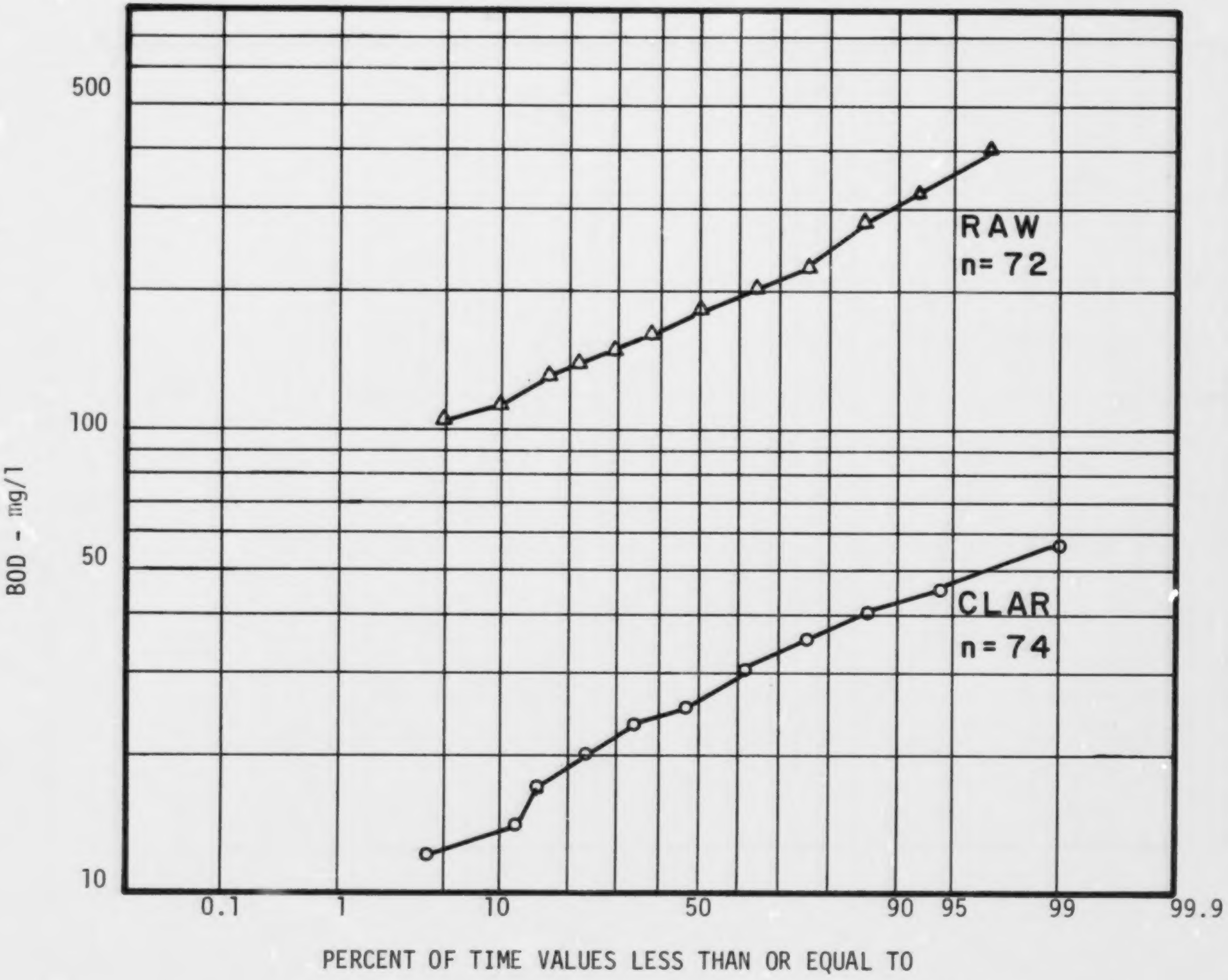


Figure 31. Frequency distributions for BOD data - 6/7/75 through 12/5/75

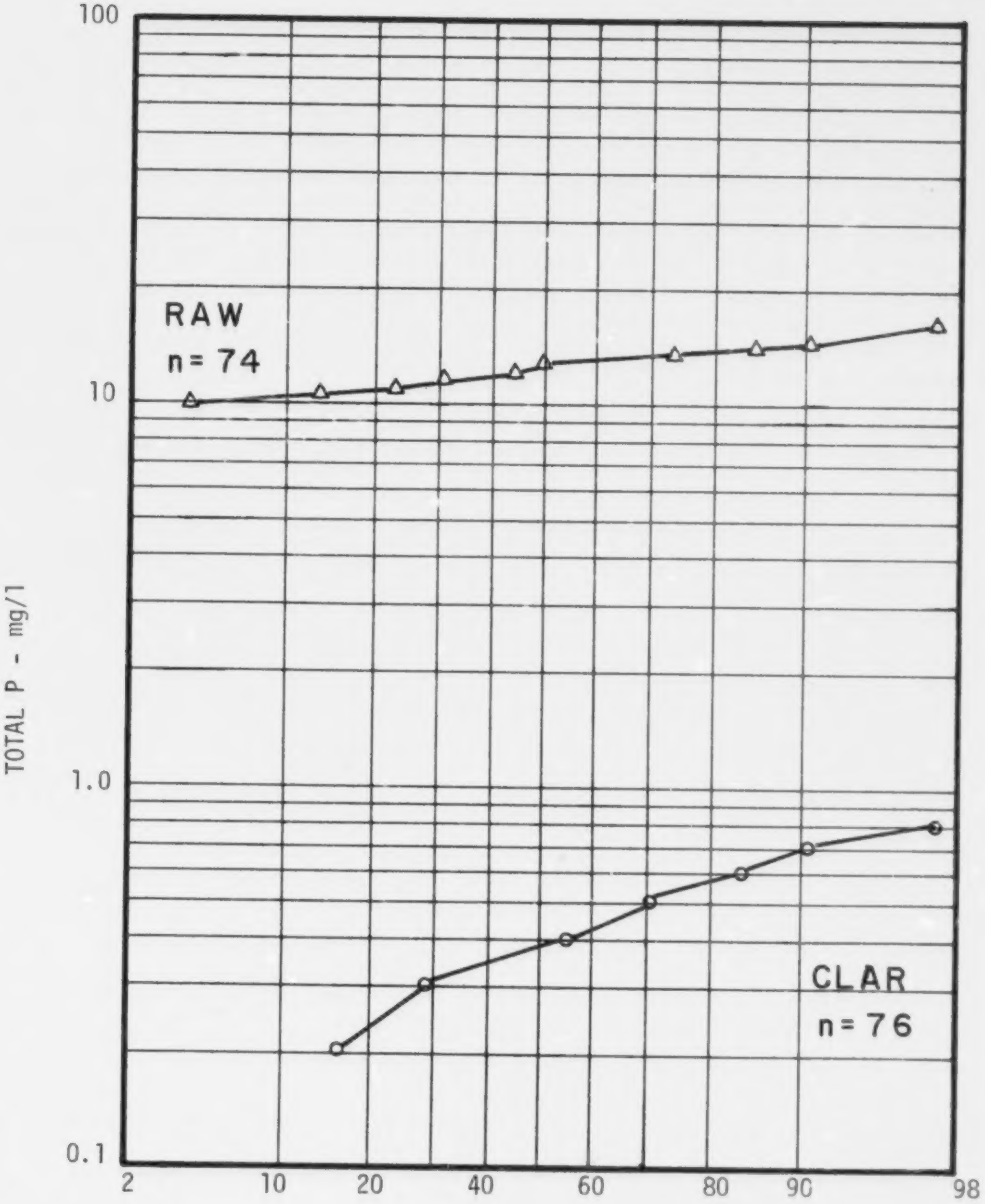


Figure 32. Frequency distributions for total P data - 6/7/75 through 12/5/75.

c. pH 9.5

For the 18-month period December 1975 through May 1977 the average of 11,939 hourly pH values was 9.64. The clarifier hydraulic loading averaged 0.49 gpm/sq ft and chemical feed rates averaged 336 mg/l lime [CaO] and 19 mg/l Fe. Based on chemical inventories conducted every 4 weeks the chemical feed rates varied from 275 to 391 mg/l for CaO and 14.5 to 22 mg/l for Fe. The $FeCl_3$ dose was varied as necessary to maintain the total phosphorus concentration at or below 1 mg/l. The average performance characteristics are presented in TABLE 39. These data demonstrate a measurable performance degradation when compared to the data obtained at pH 10.5. The frequency distributions presented in Figures 30 to 35 illustrate that the treatment variability was also much greater during the period the pH set point was 9.6. The ranges for 90% of the observations (5th through 95th percentile) for the three parameters are presented in TABLE 40.

TABLE 39. CLARIFIER PERFORMANCE CHARACTERISTICS AT pH 9.6

Parameter	Average Concentrations - mg/l Raw	Average Concentrations - mg/l Clar	% Removed *
Suspended solids	227	22	88
Turbidity (NTU)	-	10	-
BOD - total	215	44	75
BOD - soluble	62	26	50
COD - total	494	96	76
COD - soluble	164	66	51
Total P	11.4	1.1	88
Total soluble P	8.2	0.5	92
Kjeldahl-N	43	34	4.5

* based on mass flow rates in raw and clarifier flows

During this period the sludge wasting rate averaged approximately 4200 gpd at an average total solids concentration of 8.15%. Based on these vlaues the solids wasting rate averaged approximately 11,400 lbs per million gallons raw wastewater treated.

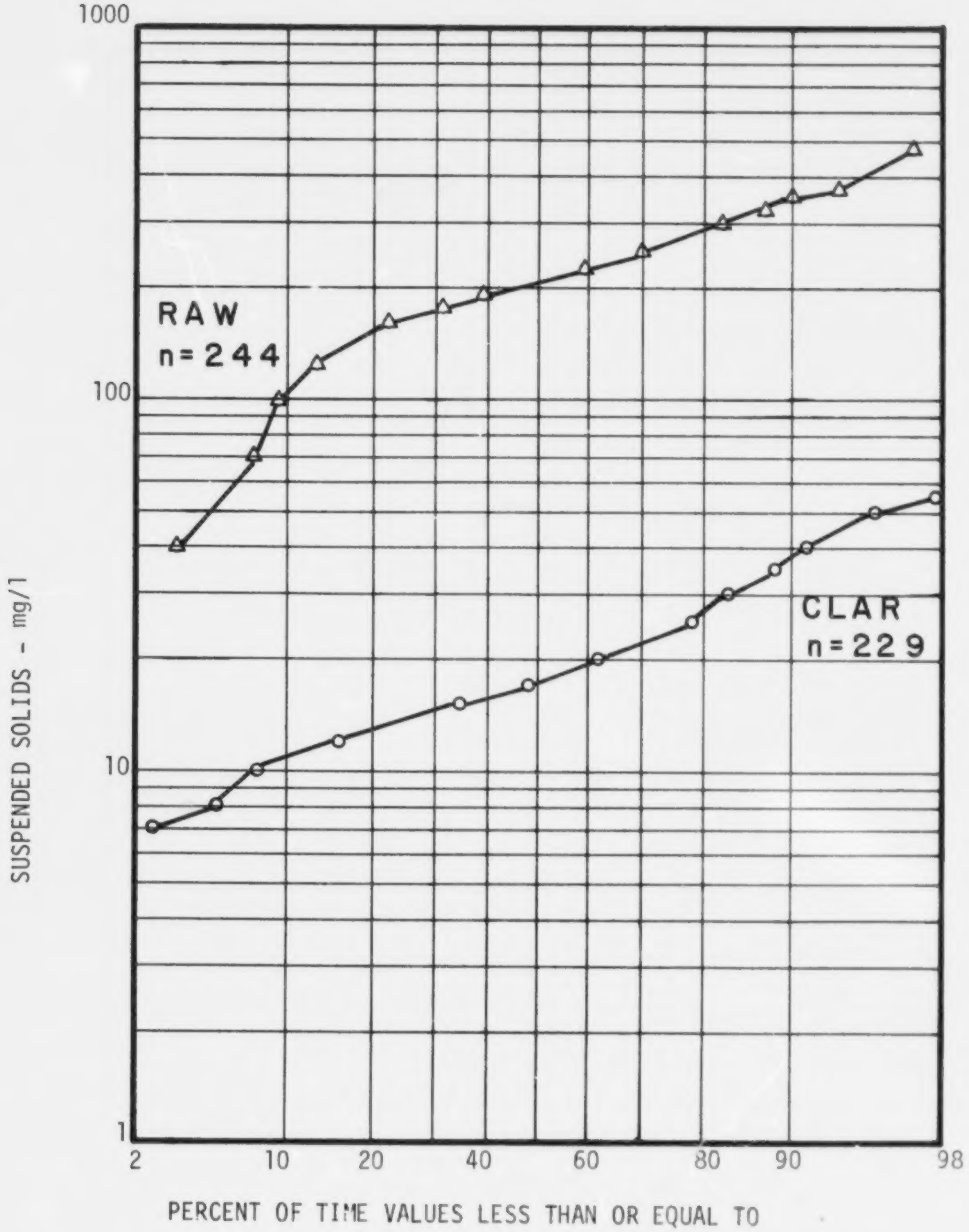


Figure 33. Frequency distributions for Suspended Solids data - 12/6/75 through 6/6/77.

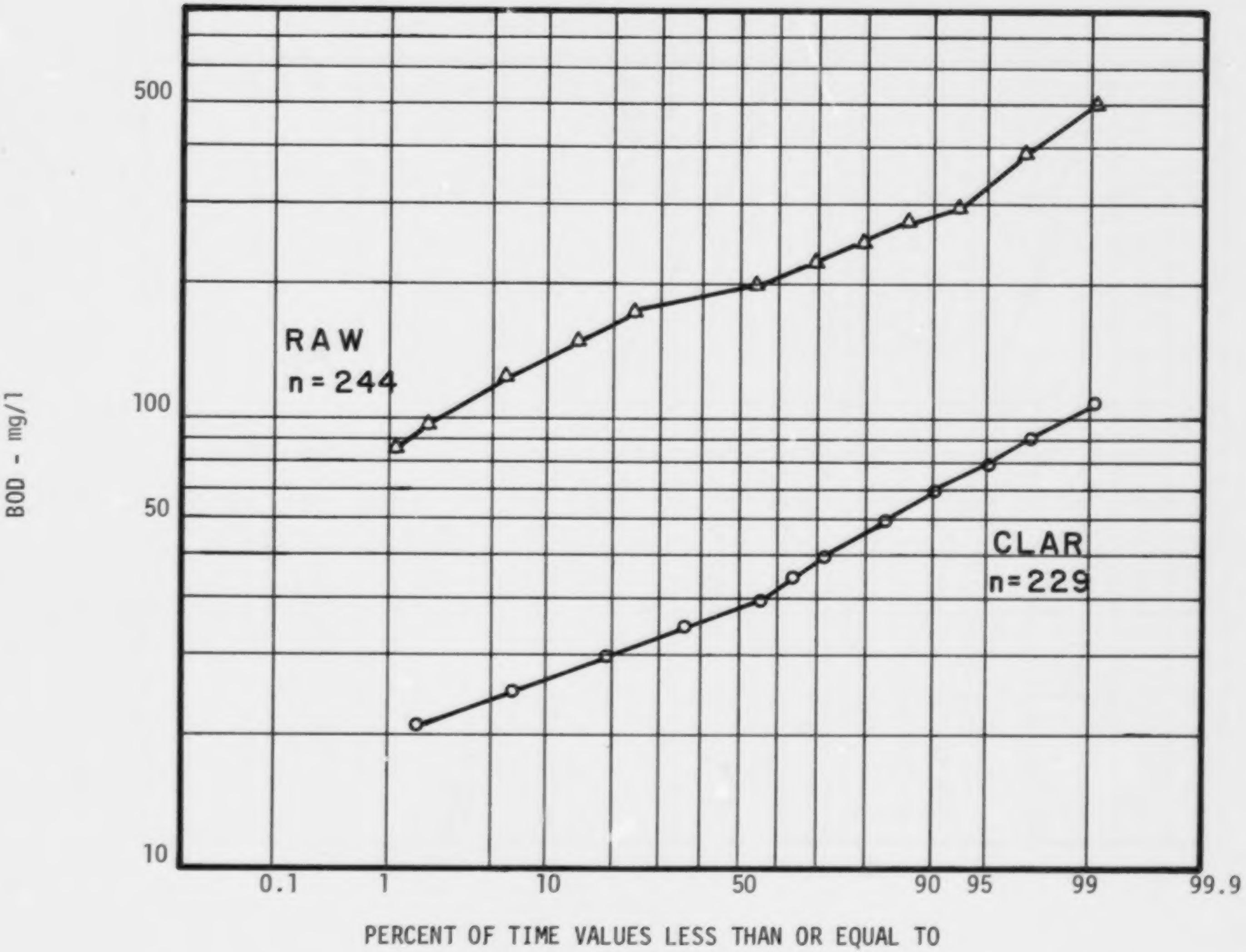


Figure 34. Frequency distributions for BOD data - 12/6/75 through 6/6/77.

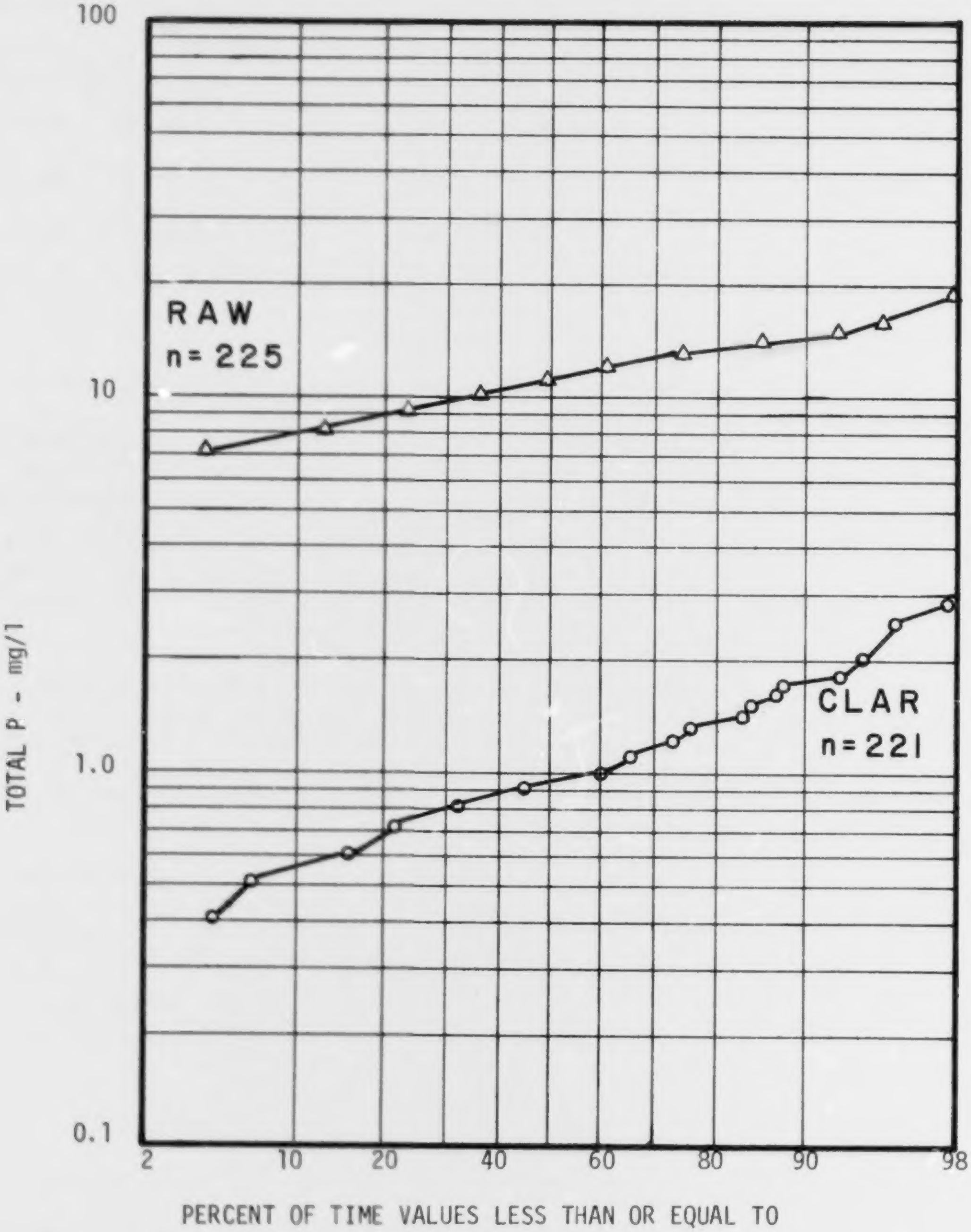


Figure 35. Frequency distributions for total P data - 12/6/75 through 6/6/77.

TABLE 40. RANGE OF CLARIFIER EFFLUENT PARAMETERS 5TH THROUGH 95TH PERCENTILE

Parameter	Clarifier pH 10.5	 9.6
Suspended solids	3 to 22 mg/l	8 to 47 mg/l
BOD - total	12 to 46 mg/l	24 to 78 mg/l
Total P	0.2 to 0.75 mg/l	0.4 to 2.2 mg/l

4. Mixing Modifications

Several attempts were made to increase the average velocity gradient in the flocculation zone and thereby increase removals of colloidal and supracolloidal particles. In December 1975 a small centrifugal pump was mounted just below the weir box and clarifier effluent was recycled to the flocculation zone. The flow discharged through a 1/2 in. horizontal pipe tangentially with respect to the skirt dividing the flocculation and settling zones. The Q vs H data supplied with the pump indicated a discharge in the range of 30 to 40 gpm however visual observations indicated the actual discharge was considerably less. The discharge did cause a slight rotational motion in the flocculation zone but had no measurable effect on clarifier performance.

Starting in January, 1976 air was injected into the flocculation zone along with the pumped recirculation. The air was discharged through three 1/2 in. pipes located just inside the skirt and approximately 4 ft below the water surface. Air mixing did not have an observable effect on clarifier performance however the concentration of the blowdown sludge increased by over 2% solids within two days after mixing was initiated.

This mixing scheme was used for approximately four months. During that time the air flow rate was adjusted frequently because of what appeared to be a gradual plugging of the discharge pipes. On two occasions the turbulence generated inside the skirt was sufficient to generate a flow through the scum collector box and to the sludge holding tank. On one of these occasions the sludge holding tank overflowed. Because of this problem and the fact that the increased mixing did not measurably affect clarifier performance the mixing modifications were terminated.

5. Costs

a. Chemicals

Clarification was one of two unit processes in the treatment train which required significant expenditures for chemicals. During the summer of 1977 the delivered unit costs for the chemicals required were as follows:

Lime	- \$0.0329 per lb CaO
Ferric Chloride	- \$0.269 per lb Fe
Sulfuric Acid	- \$0.079 per lb S

The unit chemical costs for treatment are summarized in TABLE 41. The chemical cost during operation at pH 10.5 was approximately 4.5% greater than during operation at pH 9.6. Polymer costs are not included because it was found that polymer addition did not improve clarifier performance and was therefore terminated after 5 months use.

TABLE 41. CLARIFICATION CHEMICAL COSTS - $\$/10^6$ GAL

	Clarifier pH	
	9.6	10.6
Lime	92.09	122.79
Ferric chloride	42.38	27.74
Sulfuric acid	38.36	40.04
Total	172.83	190.57

b. Labor

The cost accounting system that was utilized divided labor into two categories, direct and maintenance. Direct labor was defined as time that the operating staff spend in the appropriate cost center. Typical tasks for cost center 30 - (Clarification and Phosphorus Removal - TABLE 12) included: locating the sludge/water interface and checking flow to pH monitors on a semi-hourly basis, checking chemical feed pumps and making visual inspections of clarifier at least once per 8 hr shift, making batches of lime slurry and transferring to slurry storage tank as required, lubricating all drives as per suppliers' recommendations and draining and cleaning the clarifier as required. Maintenance labor was defined as time spent by the pipefitter and the electrician performing the major nonroutine maintenance tasks such as repair of chemical feed pumps and controllers, modification of chemical feed lines, repair of motors and drives and cleaning the clarifier influent piping of debris.

The direct labor requirements for clarification totaled 5816 hours and averaged 26.8 hr per million gal. The maintenance labor totaled 936 hours yielding an average of 4.3 hr per million gal. Total labor cost, including fringe benefits, averaged \$265 per million gal.

c. Supplies

The total expenditures for materials and supplies used for maintaining equipment plus the cost of contract maintenance totaled \$8831. The unit treatment cost was approximately \$40.71 per million gal. This cost includes repair and replacement of several circuit boards in the chemical feed pump controllers located in the plant's main control panel.

d. Power

The power consumption data for the drives associated with the clarifier are summarized in TABLE 42. The values presented are based on wattmeter readings for those motors which ran 24 hr per day and on watt-hour meter and on-time clock readings for those which ran intermittently. The average unit cost of power during the 2 year demonstration period was \$0.022 per KWH.

TABLE 42. CLARIFICATION POWER COSTS

Unit	KWH/10^6 gal	\$/10^6 gal
Lime slurry mixer	40	0.88
Lime slurry transfer pump	3	0.07
Lime slurry storage tank mixer	225	4.95
Lime slurry feed pump	61	1.34
Clarifier mixer	25	0.55

B. FILTRATION

1. Operating Procedures

Shortly after the plant was put into service it became obvious that the backwash storage pit was much too small to allow for storage of all recycle flows and discharge to maximize flow equalization. Starting early in 1974 all routine backwashing of filters and carbon columns was accomplished on a time basis. The normal backwash procedure consisted of a 5-min air scour at a rate of approximately 5 cfm/sq ft. Each filter in service was backwashed once per day according to the schedule and more often as required. During the entire two year period the number of standard backwash cycles averaged 1.25 per day for the first stage filters. The routine backwash requirements averaged 2.5% and 2.3% of the treated flow for the first and second stage filters respectively.

On several occasions it was found that the standard backwash procedure did not completely cleanse the filter as evidenced by only a partial reduction of head loss. On these occasions the filter was taken out of service and a special backwash procedure used. This procedure consisted of soaking the filter media in a 0.2N NaCl solution or a chlorine solution (50 lb HTH) and subjecting the media to a multiple air scour. The length of the air scour was based on the operator's judgement. The media was then backwashed at a rate of 16 gpm/sq ft for 5 to 10 min or until the backwash water was clean based on the operator's visual inspection. A total of 37 special backwash cycles were conducted, 16 on the first stage filters and 21 on the second stage filters.

The total plant recycle flow associated with filter backwashing averaged 14,800 gpd or approximately 6.1% of the plant influent flow.

2. Removal Efficiencies

The suspended solids and BOD loadings and removal efficiencies for the first stage filters are summarized in TABLE 43. The filters substantially reduced both the suspended solids and BOD loading to the carbon columns.

TABLE 43. FIRST STAGE FILTER PERFORMANCE CHARACTERISTICS

Parameter	Period		
	Clar pH = 10.5	Clar pH = 9.5	2 Year
Suspended solids applied - mg/l	12	22	19
- psf/day	0.28	0.55	0.47
Suspended solids discharge - mg/l	6	10	10
Suspended solids - % R	50	54	47
BOD applied - mg/l	28	44	41
- psf/day	.66	1.11	1.02
BOD discharged - mg/l	19	34	31
BOD - % R	32	23	24
Soluble BOD applied - mg/l	19	26	24
Soluble BOD discharged - mg/l	20	26	26

During the six month period that the clarifier was operated at pH 10.5, the suspended solids and BOD concentrations discharged from the first stage filters (when in service) were well within the limits defined as secondary treatment. For this period all of the insoluble BOD was removed by filtration.

The suspended solids removal efficiency did not appear to be a function of the solids loading rate; however, the removal of insoluble BOD decreased from 100% to 55% when the loading rate was increased.

The suspended solids loading of the second stage filters was extremely low even during treatment Modes I, II, and III when the carbon columns were operated in the upflow mode. Because of the position in the treatment scheme only relatively small diameter solids were applied to the second stage filters and the removal efficiency averaged approximately 30% - TABLES 30, 31, 32, 33. During treatment Modes I through IV the removal of insoluble BOD varied from 67% to 100% and averaged approximately 75%.

3. Costs

The costs associated with operating and maintaining the two sets of dual media filters are summarized in TABLE 44. The operator's tasks associated with the filters included: recording head loss of filter at end of service cycle, initiating backwash sequence, monitoring position of valves from main control panel and manually assisting malfunctioning valve positioners as required, recording head loss of clean filter at start of service cycle, adjusting sequence controller to provide multiple air scour as required. The operators also conducted the special backwashing procedures with salt or chlorine. The maintenance labor (pipefitter and electrician) and materials and supplies costs were primarily related to the repair and replacement of valves and operators. The power cost was limited to the cost of electricity to operate the blowers which provided the air for air scour. Although both sets of filters operated under conditions of gravity flow the necessary head was provided by the raw sewage pumps and the carbon column pumps. These pumping costs were absorbed by cost centers 20 and 50 respectively - TABLE 12.

TABLE 44. FILTRATION COSTS

	First Stage Filters		Second Stage Filters	
Category	Quantity	$	Quantity	$
Operating labor - hr	1,486	10,187	1,184	8,004
Maintenance labor - hr	340	2,976	346	2,940
Benefits	-	2,632	-	2,189
Materials and Supplies	-	1,033	-	869
Electricity - KWH	450	10	325	7

Based on total treated flows of 208 mil gal and 171 mil gal the treatment costs averaged $81 per mil gal and $82 per mil gal for the first and second stage filters respectively.

C. CARBON ADSORPTION

1. Operating Procedures

a. Service Time

During the 24-month demonstration project the carbon columns in each of the two treatment trains were used for extended periods. The length of time in service for an individual column was primarily based on plant effluent quality considerations. As the adsorption capacity of the carbon beds was utilized the organic content (BOD, COD, TOC) of the effluent gradually increased. When the treatment goal of 15 mg/l BOD was exceeded for 1 to 2 weeks the carbon column configuration was generally changed. The column configurations and their service times are summarized in TABLE 45.

The columns were operated in the upflow mode for approximately 6 months. During that time and previous upflow periods plugging problems were encountered with the upflow distribution header and laterals - see Figure 6. Carbon would migrate into and through the laterals and header. On November 26, 1975 column number 4 would not take flow. The header and laterals were found to be completely filled with compacted carbon. No further attempts at upflow were made.

b. Backwash

During the 6-month period the columns were operated in the upflow mode each column was backwashed once per week. The backwash procedure was as follows:

function	time
drain to air scour level	8 min
stop drain and vent	0.5 min
air scour	30 min
settle bed	5 min
backwash	15 min

The air scour rate was approximately 5 cfm/sq ft and the backwash rate approximately 12 gpm/sq ft. During the Mode II configuration, first stage filters out of service, slime accumulated in filtered water storage tank no. 1. The storage tank was readily cleaned with a salt solution. On September 17. 1975 both columns 5 and 6 were given a special backwash with 800 lbs NaCl added to each column just prior to air scour. Visual inspection of the backwash water indicated that this technique helped to remove debris which had accumulated in the carbon beds. The removal of soluble organics did increase after the special backwash but it was not repeated because the Mode II configuration was terminated shortly thereafter.

TABLE 45. CARBON SERVICE CYCLES

Service from		Column No.		Flow	
Date	Week No.	Primary	Secondary	Mode	Comments
June 7, 1975	1	4	5	up	Columns went in service March 20 and May 5, 1975
Aug. 11, 1975	10	5	6	up	Column 6 freshly regenerated
Nov. 21, 1975	24	4	-	up	Column 4 freshly regenerated
Nov. 26, 1975	25	4	-	down	Upflow header and laterals plugged with carbon
Dec. 31, 1975	30	4	5	down	Column 5 virgin carbon
Feb. 17, 1976	37	4,2	5,3	down	2 treatment trains in service because of high raw wastewater flows
Feb. 26, 1976	38	2	3	down	Maintenance on train 2
March 2, 1976	39	4	5	down	
May 18, 1976	50	5	6	down	Column 6 freshly regenerated
Aug. 25, 1976	64	1	2	down	Columns 1 and 2 freshly regenerated
Nov. 23, 1976	77	2	3	down	Column 3 freshly regenerated
Dec. 10, 1976	79	5	6	down	Maintenance on train 1, columns 5 and 6 freshly regenerated
Dec. 18, 1976	81	3	1	down	Columns 3 and 1 freshly regenerated
March 1, 1977	91	1	2	down	Column 2 freshly regenerated

During most of the approximately 18-month period the columns were operated in the downflow mode they were backwashed on a daily basis. Even with daily backwash, hydrogen sulfide generation was encountered until sodium nitrate was fed to the primary column influent line. Soon after daily backwashing was initiated observable quantities of activated carbon were found in the backwash pit. It was determined that the drain down step was not repeatable and that carbon could be blown out of the column and into the piping system during the air scour. A manual drain-down scheme was instituted to eliminate this possibility; however, carbon still accumulated in the backwash pit. The backwash rate was subsequently lowered to 8 gpm/sq ft. No significant quantities of carbon were found in samples of backwash water; however, significant quantities of carbon accumulated in the backwash pit during the remainder of the project.

c. Hydrogen Sulfide Control

During the first two years of operation (1974 and 1975) H_2S generation in the carbon beds caused odor and corrosion problems in the plant. A sampling program conducted in October 1974 indicated that the H_2S concentration in the plant ambient air ranged from 0.15 to 0.5 ppm. The minimum concentration at which the characteristic odor is preceptible is reported to be 0.13 ppm. (24) During the cold weather months when the plant doors were closed and outside air did not circulate freely the concentration were considerably higher based on odor levels. The sulfate reducing bacteria which are anaerobes appeared to thrive in the carbon beds.

Directo et al. (25) demonstrated that hydrogen sulfide generation can be controlled by providing NO_3 as an alternate hydrogen acceptor. They found that 5.4 mg/l NO_3-N in the carbon column feed eliminated H_2S production. The polymer feed system (polymer feed terminated November 16, 1975) was modified to allow the addition of sodium nitrate solution to the carbon column feed stream on a flow proportional basis. Nitrate feed was initiated on January 20, 1976 and remained in service throughout the remainder of the demonstration period. The hydrogen sulfide concentration of the GCC sample was determined on a daily basis when its characteristic odor was detected in the plant and these data are summarized in Figure 36.

Column 4 was put into service late in November 1975 (TABLE 45) and by mid January 1976 the H_2S level was as high as 12 mg/l. For the period January 20, 1976 through March 13, 1976 the carbon columns were backwashed on a daily basis and a sodium nitrate dose of approximately 7 mg/l N was maintained in the carbon column feed stream. The H_2S-S level averaged 0.5 mg/l. For the two week period starting March 14 the nitrate dose was reduced to 3.5 mg/l N and the average H_2S-S level increased to 4 mg/l. For the period April 23 through May 13, 1976 backwashing was conducted on alternate days (primary column one day, secondary column the following day) and the nitrate addition averaged 4.2 mg/l N. The H_2S-S level increased substantially and averaged 6.5 mg/l during the first two weeks of May. This average is based on only four analyses; however, odor levels in the plant were significantly higher than in previous weeks.

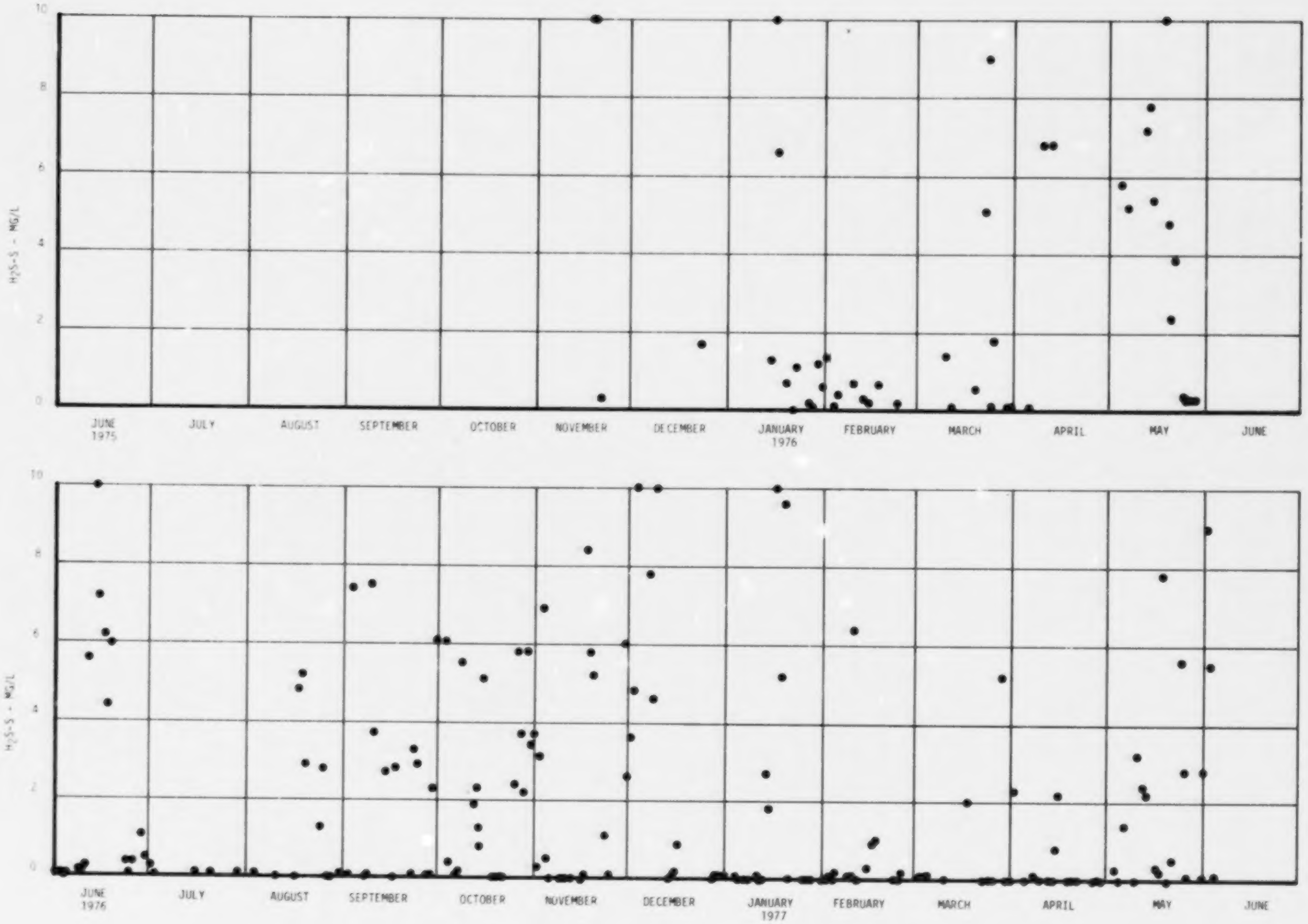


Figure 36. Hydrogen sulfide concentration of GCC sample.

During the 12-month period June 1976 through May 1977 the carbon columns were backwashed on a daily basis and the set point for NO_3-N addition was 5 mg/l. The actual dose averaged 4.8 mg/l NO_3-N based on metered carbon column flows and recorded quantities of $NaNO_3$ used. During that period the H_2S-S concentration in the GCC sample stream varied from 0 to 16 mg/l and averaged 1.9 mg/l.

The variable efficiency of sulfide control demonstrated in Figure 36 was related to three factors: failure of the chemical feed pumps, increased organic load, and operation of the carbon regeneration furnace. The most common pump failures were caused by plugging of the filter screen and air accumulation in the diaphragm and check valves. As soon as a pump failure was identified the bleed valve was opened and the filter screen cleaned. Because of its location it was not convenient to check the liquid level in the sodium nitrate solution tank. As a result pump failures could have occurred for up to 12 hours prior to detection. On several occasions when the organic content of the clarifier effluent was significantly higher than normal the H_2S levels increased. This was most likely related to an increased oxygen demand exerted by the biomass in the carbon beds.

During November of 1976 it was observed that the H_2S level in the GCC sample stream increased substantially during the period the carbon regeneration furnace was operated. The organic content of the carbon column influent stream was normal and the nitrate feed pumps were functioning. When the carbon column influent was sampled the NO_3-N concentration was found to be zero. The measured NO_3-N concentration of the sodium nitrate feed solution was found to be 4800 mg/l. When one volume of the feed solution was mixed with 99 volumes of the carbon column influent water the measured NO_3-N concentration was found to be 0.8 mg/l.

When the carbon furnace was operated, spent scrubber water and quench water were discharged to the first stage filter influent line. These recycled flows apparently contained a significant quantity of reducing agents that reacted with the nitrate ion. It was found that this recycle stream had a five minute chlorine demand of 41 mg/l. In order to eliminate the reducing agents the clarifier effluent was chlorinated heavily during subsequent carbon regenerations.

2. Removal Efficiencies

a. General

The removal efficiencies obtained with activated carbon treatment are summarized in TABLE 46. For the entire two year period the overall average removals of soluble organics were in the range of 53 to 55% depending on the parameter. Likewise overall average removals of total organics were in the range of 42 to 52%. The data presented in TABLES 29 through 34 indicate that removals of insoluble organics were substantial when the carbon columns were operated downflow - Modes IV and V.

TABLE 46. SUMMARY OF REMOVAL EFFICIENCIES ACROSS ACTIVATED CARBON COLUMNS

Constituent	Average % Removal					
	Mode I	Mode II	Mode III	Mode IV	Mode V	Total
BOD - total	38*	52*	36*	38	59	42
BOD - soluble	44	45	30	52	67	54
COD - total	23*	54*	34*	51	61	49
COD - soluble	41	57	39	55	63	55
TOC - total	50*	47*	44*	52	56	52
TOC - soluble	47	50	33	53	58	53
Suspended solids	-40	-12	-12	36	44	30

* Removal obtained in second stage filters included when carbon columns operated upflow.

The average soluble COD loading was calculated for each of the 104 weeks of study period. The mass of soluble COD in a particular flow stream was defined as the product of the total weekly mass flow (mil lbs) and the average of the weekly analyses (usually 3). The loading calculations are summarized in Figure 37. Although there were no dramatic changes in carbon efficiency throughout the study the data can be divided into 3 groups. During the first 6 months, when the clarifier was operated at pH 10.5 and the carbon columns were operated upflow, both the rates of soluble COD application and removal were relatively low. During the following 12 month period both rates were slightly higher and during the final 6 months slightly higher still. The columns were operated in the downflow mode during the latter two periods. Although no attempt was made to regulate the service times or ultimate loadings of the individual columns it does appear that both the mode of operation (upflow vs downflow) and the application rate (concentration) did affect the removal efficiency of soluble COD.

b. Upflow

Columns No. 4 and 5 were put into service in the upflow mode early in May 1975 - TABLE 45. The upflow mode was utilized until late November 1975. During that period approximately 10,200 lbs soluble COD were removed in the carbon beds. Columns 4, 5, and 6 contained an estimated total of approximately 51,000 lbs freshly regenerated activated carbon (the bulk density of the carbon in place was estimated to be 26 lb/cu ft) yielding a carbon loading of 0.20 lb soluble COD per lb carbon. Directo et al. (25) reported soluble COD removal efficiencies on the

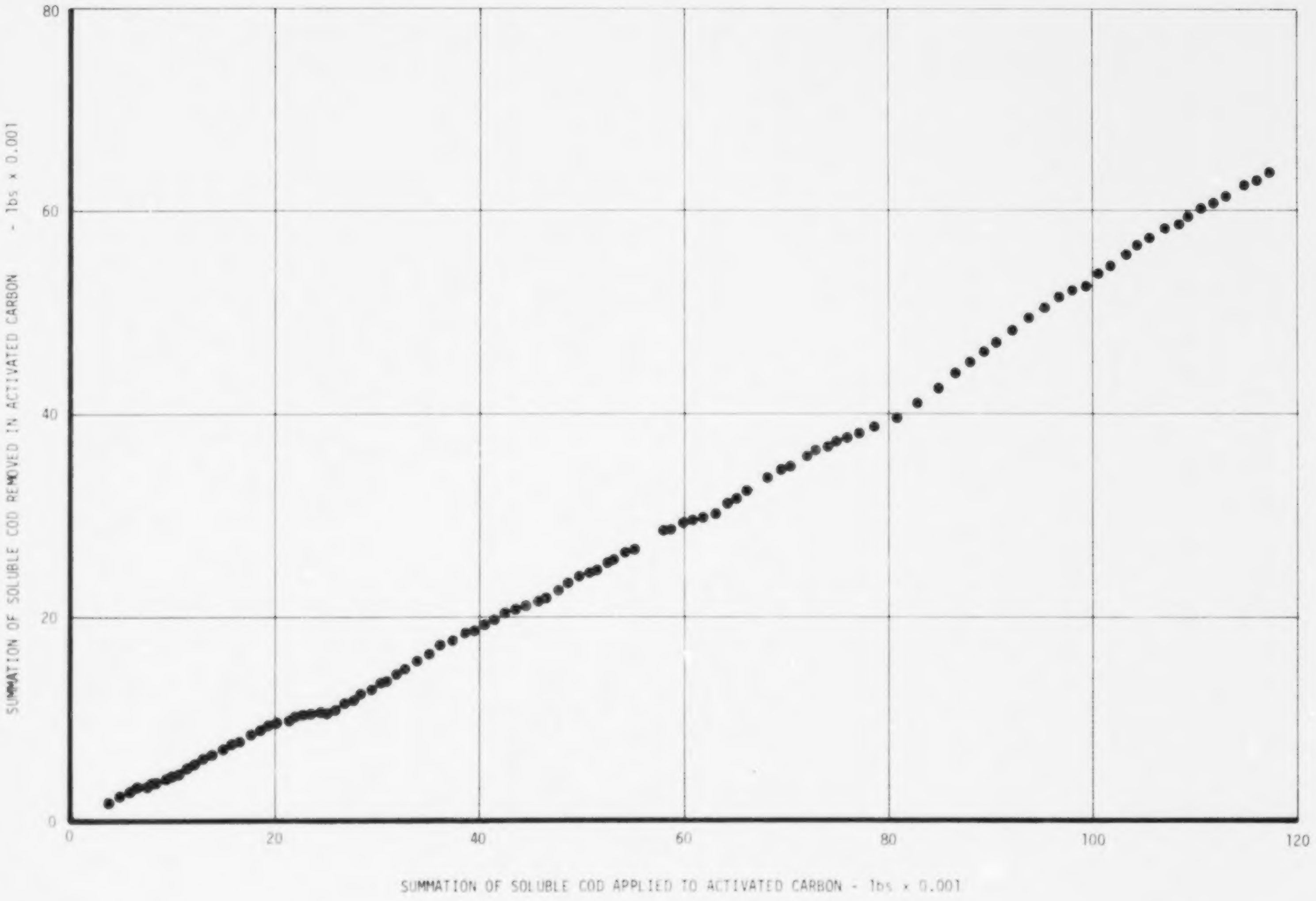


Figure 37. Summary of soluble COD application and removal rates.

order of 65% when the loading approached 1.5 lb soluble COD per lb carbon and the concentration applied averaged approximately 50 mg/l. The data presented in Figure 38 illustrate that the removal efficiency decreased significantly as the carbon loading increased. The average concentration of soluble COD applied to the carbon was slightly less than 50 mg/l. The extremely low removal efficiency of 11% for the fourth week was related to the pH control problem discussed in the previous section. Based on the measured parameters plant operation was normal for the remainder of the period. Because the GCC_1 sample stream was not collected until late in the study no attempt was made to determine the loadings and removals for the individual columns.

c. Downflow

Column 4 was put into service on November 21, 1975 in the upflow mode and switched to downflow five days later. The soluble COD removals obtained with columns 4, 5, and 6 in the downflow mode are summarized in Figure 39. The low removals obtained during the first two weeks were related to a valve malfunction. After the switch to downflow operation the DMF_1 and GCC samples indicated no organic removals; however, samples taken directly from the column sample taps indicated almost complete COD removal. This anomaly was resolved on December 9, 1975 when it was discovered that a valve on the carbon column inlet header (downflow mode) was open allowing essentially all of the flow to bypass the carbon column. The valve malfunction was not found earlier because the valve indicator lights on the main control panel showed the valve in the closed position. The valve actuator was in the closed position but the key in the mechanical connection sheared and the valve remained open. Unfortunately the limit switch used to indicate position was mounted on the actuator. In addition, this valve, along with most others on the carbon columns, was mounted above the column approximately 30 ft above the floor and was not readily accessible for inspection.

The removal efficiency dropped below 30% on two other occasions. On both occasions the pH of the carbon column feed stream was elevated by the return of zeolite column rinse water to filtered water storage tank No. 2. This problem is discussed more fully in the following section.

During the 40-week period that columns 4, 5, and 6 were in service the loading averaged 0.38 lb soluble COD per lb carbon. The data presented in Figure 39 again indicate that the removal efficiency was a function of loading.

Sometime after column 4 was taken out of service it was discovered that approximately 85% of the column's initial charge of carbon had been lost. It was never determined over what time period the losses took place but it is assumed that they occurred during column backwash. Because of this loss the actual loading was somewhat higher than the calculated value of 0.38 lb per lb.

Columns 1, 2, and 3 were utilized for the last 39 weeks of the study. The soluble COD removal data for this period are summarized in Figure 40. The rate at which soluble COD was applied to the carbon columns was substantially greater during this last period because of increased flow and concentration. When compared to the initial 6-month period the flow increased by approximately 9%

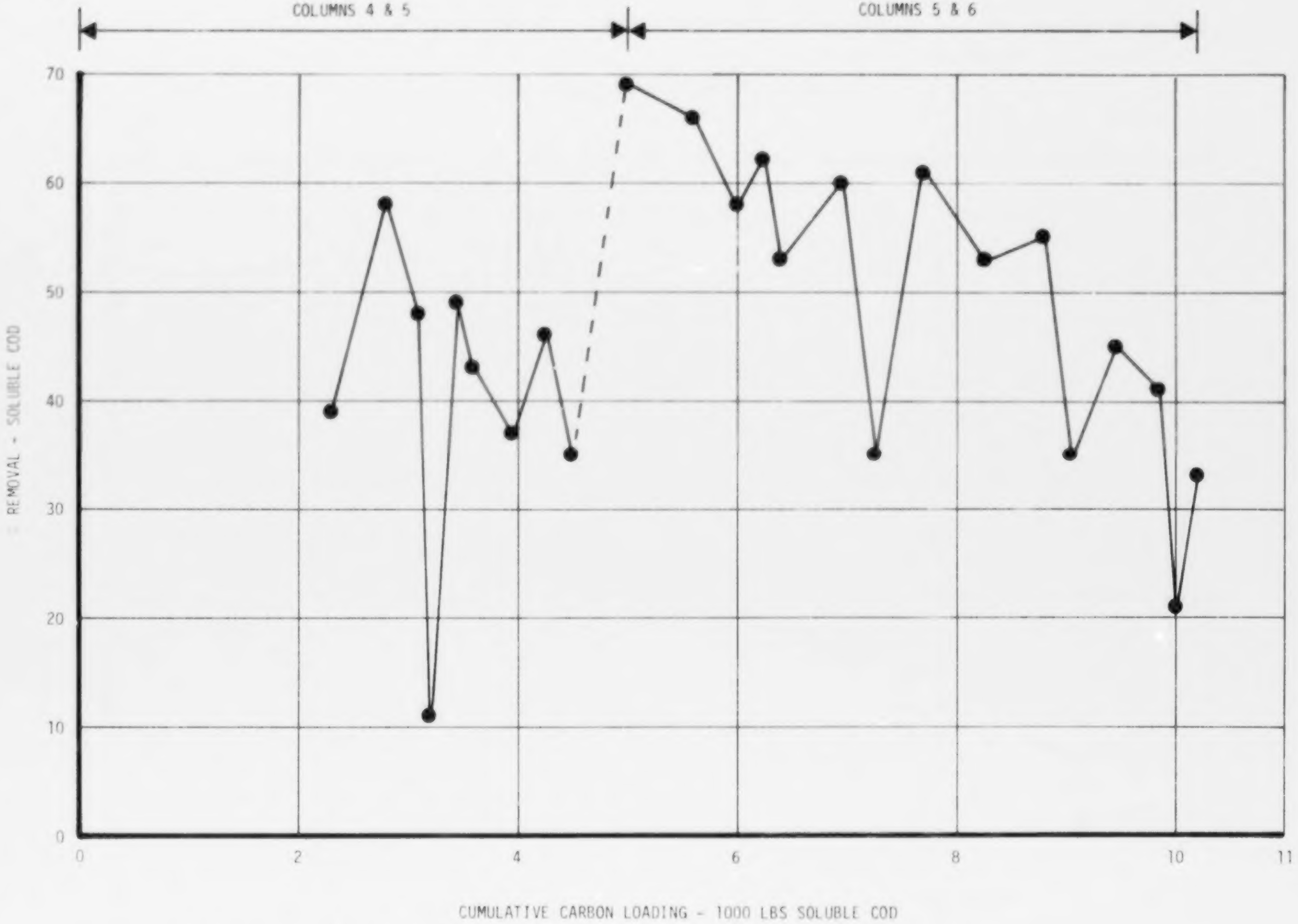


Figure 38. Soluble COD removal across carbon columns 4, 5 and 6 during upflow operation.

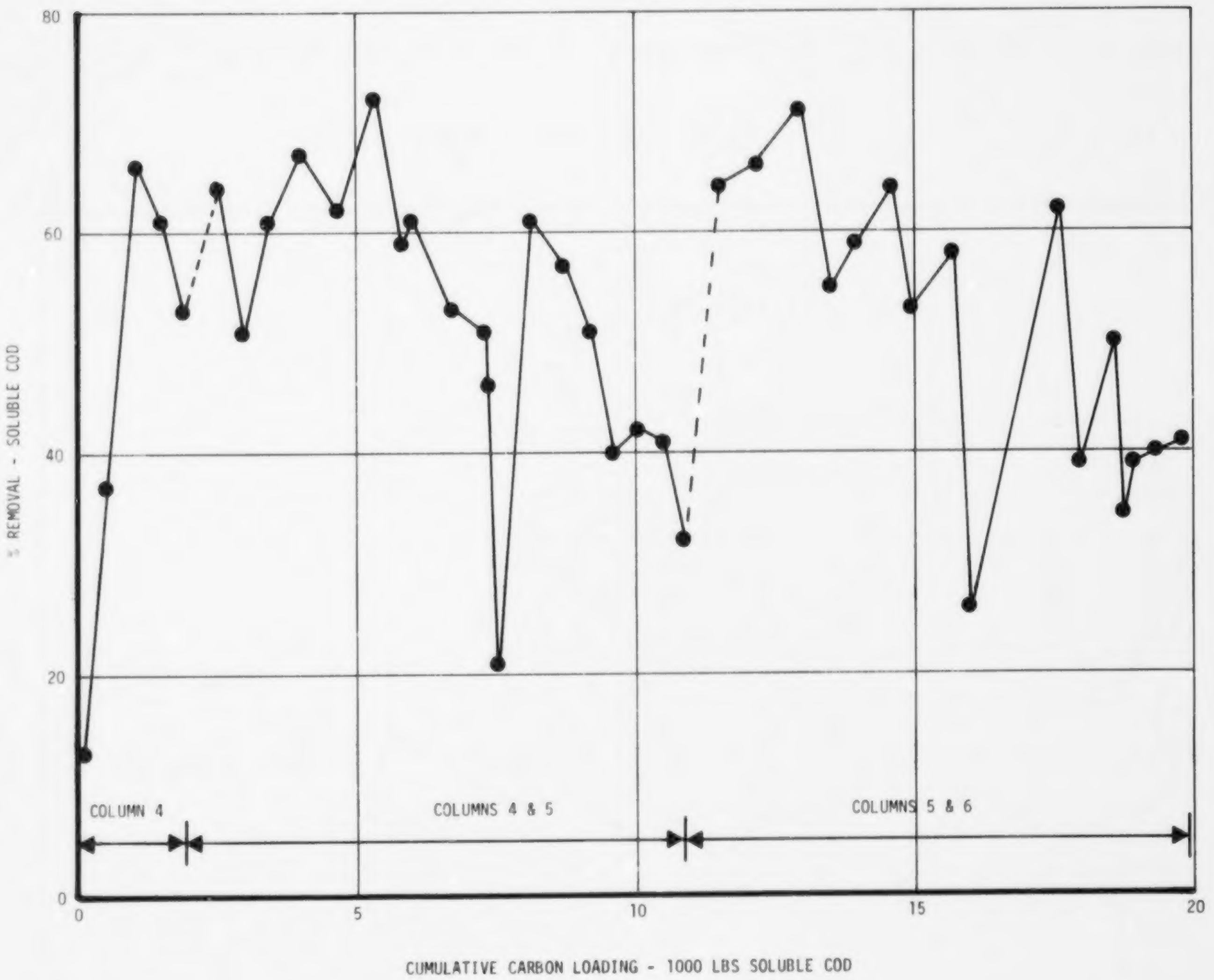


Figure 39. Soluble COD removal across carbon columns 4, 5 and 6 during downflow operation.

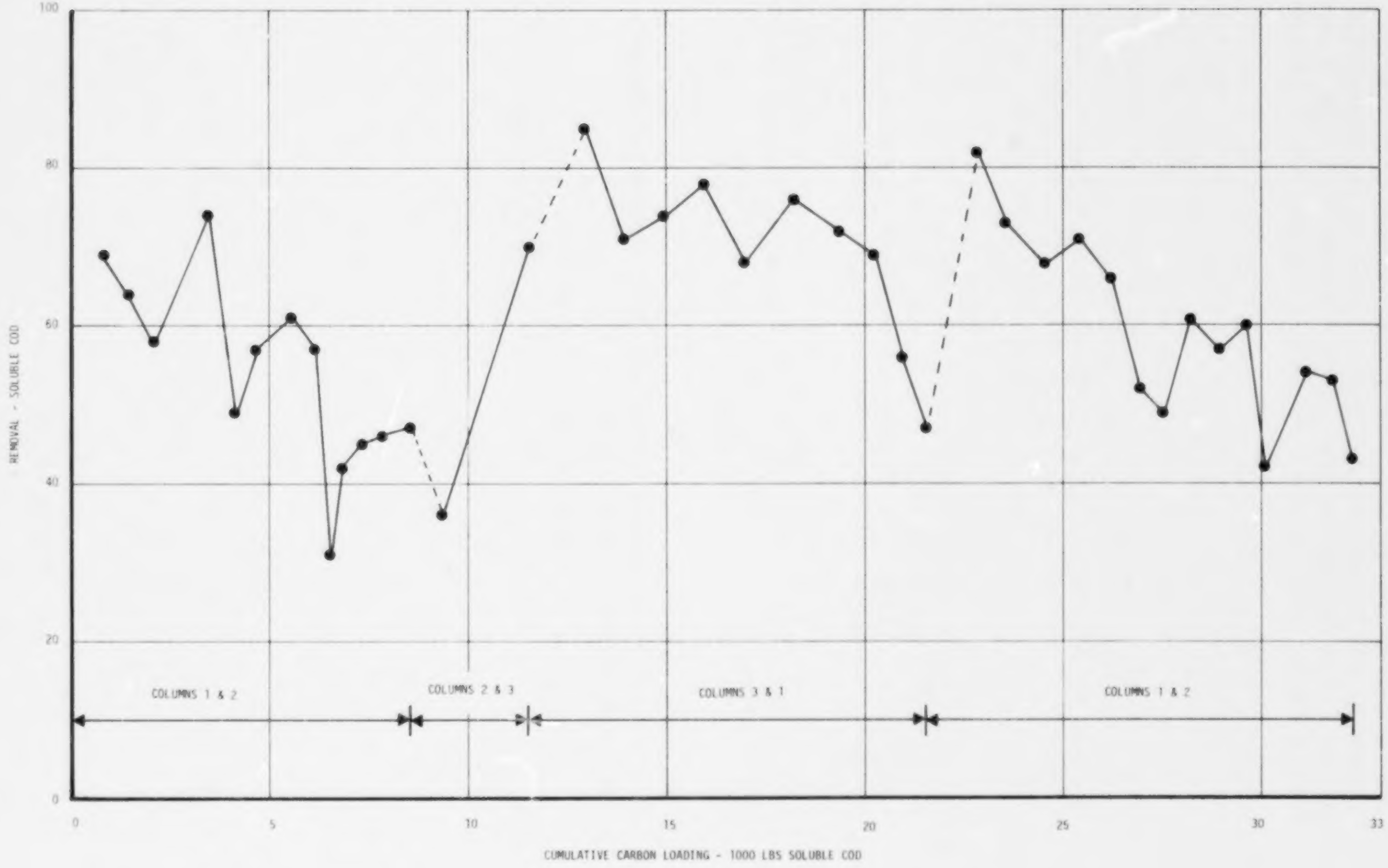


Figure 40. Soluble COD removal across carbon columns 1, 2 and 3 during downflow operation.

and the soluble COD concentration by 40%. The higher removal efficiencies illustrated in Figures 39 and 40 were most likely related to the higher COD concentrations applied. For the period that carbon columns 1, 2, and 3 were in service the average carbon loading was 0.40 lb soluble COD per lb carbon.

The carbon loadings obtained during this study, 0.20 to 0.40 lb soluble COD/lb carbon were substantially lower than those obtained at pilot facilities in recent years where values in the range of 0.36 to 1.5 have been reported (25)(26)(27)(37). Because the COD test is not specific it is difficult to determine why such a wide variation exists. Although it has been demonstrated that the type of pretreatment can affect carbon performance (30) the variations are most likely related to the chemical nature of the waste stream and the physical/chemical nature of the carbon used. The degree to which pure organic compounds are adsorbed varies substantially. (31) A wastewater system is further complicated by the synergistic and inhibitory effects that may be produced by some compounds on others.(32) It is thus not unexpected to find variations in treatment efficiency from plant to plant but also from day to day within a plant. This is especially true for small plants such as Rosemount where relatively small discharges of specific classes of compounds may have a significant effect on the performance of the carbon adsorption system.

It is also possible that significant portions of the organic removals reported here and elsewhere were related to the production of biomass which further complicates attempts at data comparisons.

3. Costs

The costs associated with operating and maintaining the carbon columns of trains 1 and 2 are summarized in TABLE 47.

TABLE 47. CARBON ADSORPTION COSTS

Category	Quantity	$
Director labor - hr	1,957	13,452
Maintenance labor - hr	120	1,081
Benefits	-	2,898
Carbon - lb	24,000	11,520
Sodium nitrate - lb	41,200	5,460
Materials and supplies	-	1,813
Contracted services	-	700
Electricity - KWH	140,500	3,091

The operators' routine tasks consisted of monitoring and recording head loss across the carbon beds at least once a shift and monitoring the backwash operation. Nonroutine tasks consisted of removing carbon from the upflow distributors and headers (during that period when the columns were operated in the upflow mode) and sampling backwash water for carbon, cleaning carbon and other debris from the plant backwash pit and loading makeup carbon into the columns. Maintenance labor consisted primarily of repairing valves and operators that malfunctioned. The contracted services consisted of installation of an additional manway in column 4 to facilitate inspection of the underdrain system. The electricity cost was estimated from the run time for the carbon column pumps and the blowers used for air scour prior to backwash.

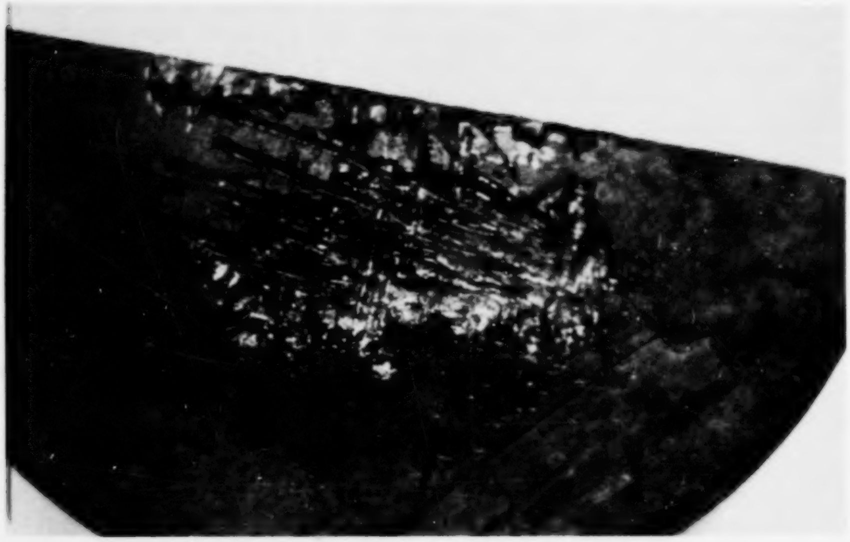
As of November 1977 at least one of the carbon columns requires considerable repair before it could be put back into service. A reliable cost estimate for the required repairs was not available at the time this report was prepared.

4. Corrosion Problems

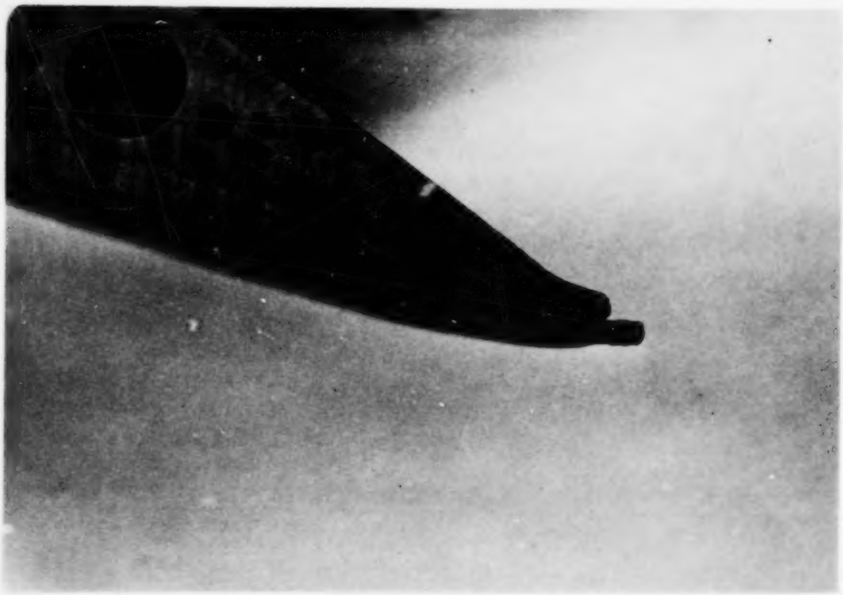
A previous section alluded to some of the problems associated with H_2S generation including aesthetic considerations and severe corrosion of exposed copper and brass fittings and electrical gear. The failure of several copper tubing air lines was attributed to this type of corrosion. (They were replaced with polyethylene tubing.) In November 1976 the process water inlet valve on one of the second stage filters (valve A, Figure 4) failed to seat properly. The valve was removed and the disc was found to be covered with a scale which appeared to be iron oxide. In the process of removing the scale the pipefitter observed that the disc metal (cast iron) was soft and could be removed with the fingernail - Figure 41A. A commercial laboratory examined the disc and found the soft black corrosion product to be approximately 1/8 in. thick - Figure 41B. The results of a chemical examination of the core material and the corrosion product are presented in TABLE 58. It appears the disc was undergoing graphitization.(28) No attempt was made to examine other unprotected metal surfaces that were contacted by process flows containing hydrogen sulfide; however, in January, 1977 the flow tube used to meter train No. 1 effluent flow was removed because of obvious malfunction. The unit, Figure 42, was made up with a steel flange, steel equalizing ring on the upstream face (with two pressure taps), brass throat with two pressure taps and a fiberglass diffuser section. Both of the pressure taps on the upstream face were covered with heavy scale and several additional holes had developed. In addition the seal between the high and low pressure chambers was corroded through.

TABLE 48. VALVE DISC ANALYSES

Parameter	% By Weight	
	Core Material	Corrosion Product
Carbon	3.21	10.20
Silicon	2.15	5.92
Manganese	0.87	1.26
Phosphorus	0.12	0.27
Sulfur	0.09	1.44

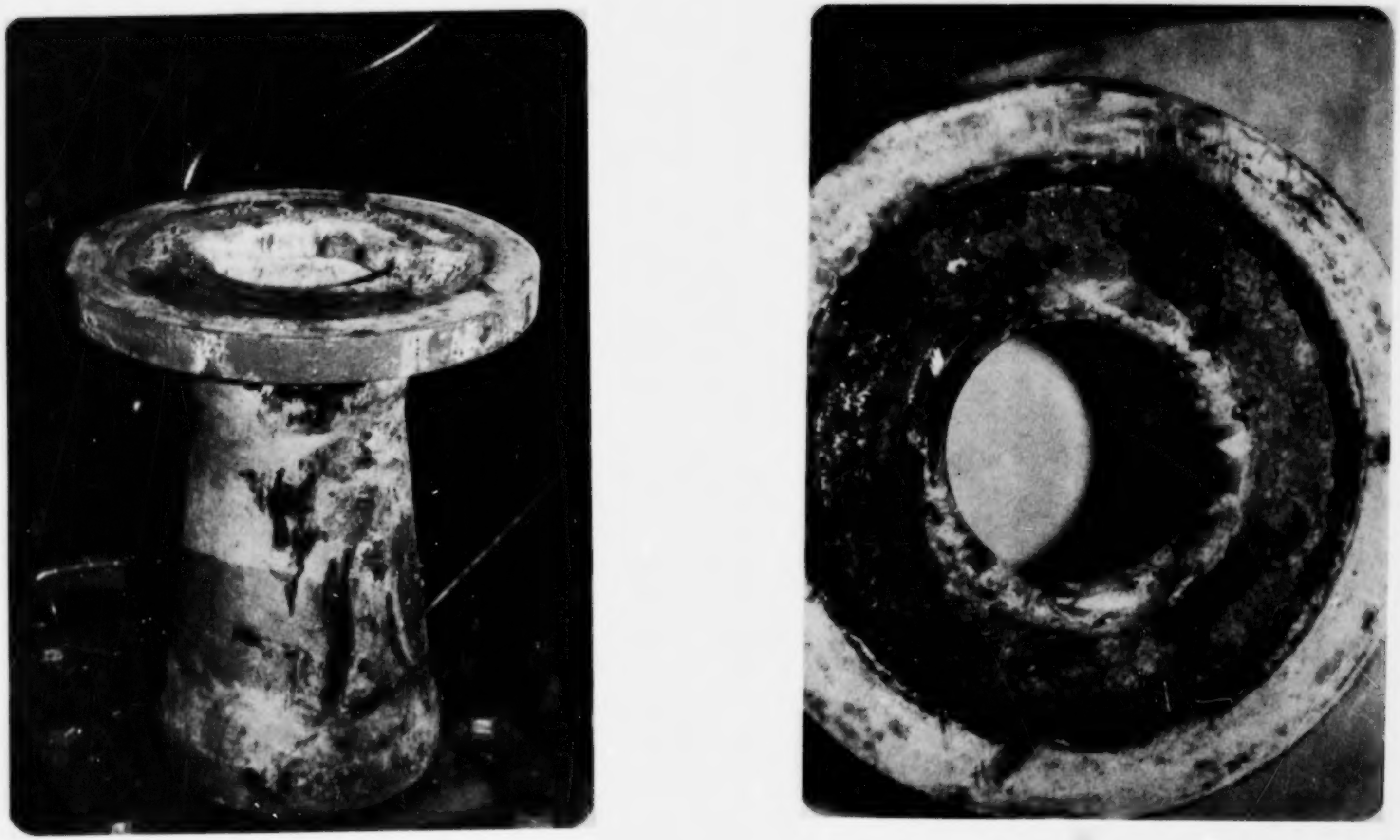


(a) Surface



(b) Section

Figure 41. Corroded valve disc.



(a) Side View

(b) Top View

Figure 42. Corroded flowmeter.

In mid September 1976 the carbon was transferred from column No. 4 to the dewatering tank - Figure 8. The operators noted that the transfer took less than half the normal time period. Based on the level in the dewatering tank it was estimated that only 2 feet of carbon were transferred from column 4 whereas its initial charge was 12.5 ft. When the lower hatch on the column was opened approximately 1 gal of carbon was found. (The lower hatch it located below the underdrain system but is positioned in such a manner that the underdrains cannot be inspected.) It was assumed that the carbon was lost through one or more underdrains that may have failed. An additional manway was installed in the column just above the underdrain level. All of the funnels were removed and the strainers inspected - Figure 43a. All were intact. It is assumed that the carbon was lost during the backwash procedure.

The initial inspection of the column interior revealed considerable corrosion damage to the steel. The tank interior was initially painted with one coat of epoxy primer and two coats of coal tar epoxy as per the construction specifications. This protective coating had been almost completely removed in the lower 12 ft of the column. The tank wall was constructed of 3/8 in. mild steel and some of the pitting reached a depth of 1/8 in. - Figure 43b, c. The interior welds were also severly corroded. As of November 1977 no decision has been made on how to proceed with the necessary repair.

5. Carbon Losses

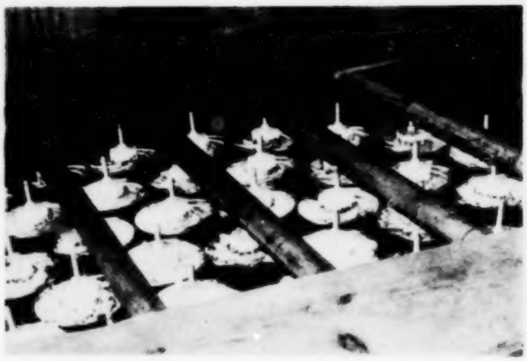
The initial carbon inventory when the plant was put into service was approximately 109,000 lbs. The carbon inventory as of October 10, 1977 was approximately 85,000 lbs. Approximately 24,000 lb carbon was purchased and added to the columns during the approximately four years of operation. Total carbon losses were approximately 48,000 lbs or 146 lbs per mil gal raw wastewater treated. During the period November 19, 1973 through October 10, 1977, 27 carbon regnerations were accomplished. A literature review indicates expected carbon losses fall in the range of 5% to 10% per transfer/regeneration cycle. (2)(6)(13)(29). The average observed loss was 1,777 lb per regeneration cycle or approximately 11.4%.

Because of the difficulties encountered, the carbon inventory was not determined on a regular basis; however, during the demonstration period the volume of each regenerated carbon batch was determined in the regenerated carbon storage column (Figure 8) prior to transfer. The losses were thus documented; however, it was not determined if the losses were related to the regeneration cycle or to the service cycle. In the case of column 4 previously discussed it was demonstrated that a major loss was related to the service cycle. It is estimated that more than 50% of the carbon loss was associated with the service cycle and the backwashing procedure specifically.

D. ION EXCHANGE AND ZEOLITE REGENERATION

1. General

Early in the investigation it was determined that the treatment goal of 1 mg/l NH_3-N would not be attainable unless significant changes were made.



(a) floor

(b) wall

(c) wall

Figure 43. Interior views of carbon column number 4.

The plant design was based on an average NH_3-N concentration of 15 mg/l which was obtained at the old treatment facility. The two-year average NH_3-N at the new facility was 35 mg/l (TABLE 14) and the zeolite columns were overloaded when the plant was put into service. At that time two alternatives were available: reduce the length of the column service cycle or modify the regeneration procedure. If the service cycle time was reduced more than one regeneration cycle would be required per day. If the columns were operated to breakthrough and then regenerated the backwashing schedule for the entire plant would have been disrupted. If two or more complete regeneration cycles were required per day the schedule would not be disrupted but the labor requirement, both operating and maintenance, for the regeneration process would have increased substantially.

Koon and Kaufman (33) demonstrated that the efficiency of regeneration with a constant brine volume was a function of the sodium concentration and the pH. Because of the ease of implementation the early modifications to the regeneration scheme at Rosemount were centered on these two parameters. The process data however indicated that the regenerant volume would have to be increased substantially in order to obtain more efficient regeneration. The latter modifications allowed the regenerant volume to be increased by a factor of up to 3.3.

2. NH_3-N Removal as a Function of Regeneration Conditions

a. Scheme 1, pH 11 - 4.6 BV - 1 N Na

For 23 days the zeolite regeneration system was operated according to the operating instructions supplied by the contractor. Two zeolite columns were in service - one primary and one secondary. At midnight the secondary column was moved to primary position, the primary column was taken off line for regeneration and a freshly regenerated column put into service in the secondary position. The regeneration conditions were as follows: 4.6 BV brine at pH 11.0 and 1.0 N Na - see TABLE 26. During the first three days of operation the removals averaged only 10% and it is assumed that some errors were made in the regeneration procedure. For the remaining operating days the column influent and effluent concentrations averaged 24 mg/l NH_3-N and 12 mg/l NH_3-N respectively.

Columns 4 and 5 were both regenerated and put into service at 0800 hr on June 26, 1975. The influent and effluents were sampled every two hours until 1200 hr on June 28, 1975. The data obtained are summarized in Figure 44. Initial leakage from both columns was high. The primary column equilibrated after 118 BV throughput and the secondary column after approximately 237 BV. A total of 73 lbs NH_3-N was removed. The two columns combined contained approximately 27,600 lbs clinoptilolite, assuming a bulk density of 46 lbs/cu ft (34). Under the test conditions the effective dynamic capacity of the zeolite was approximately 0.19 me/gm.

The data obtained on regeneration chemical requirements were not valid because of the atypical brine losses that occurred - see Section IX G.

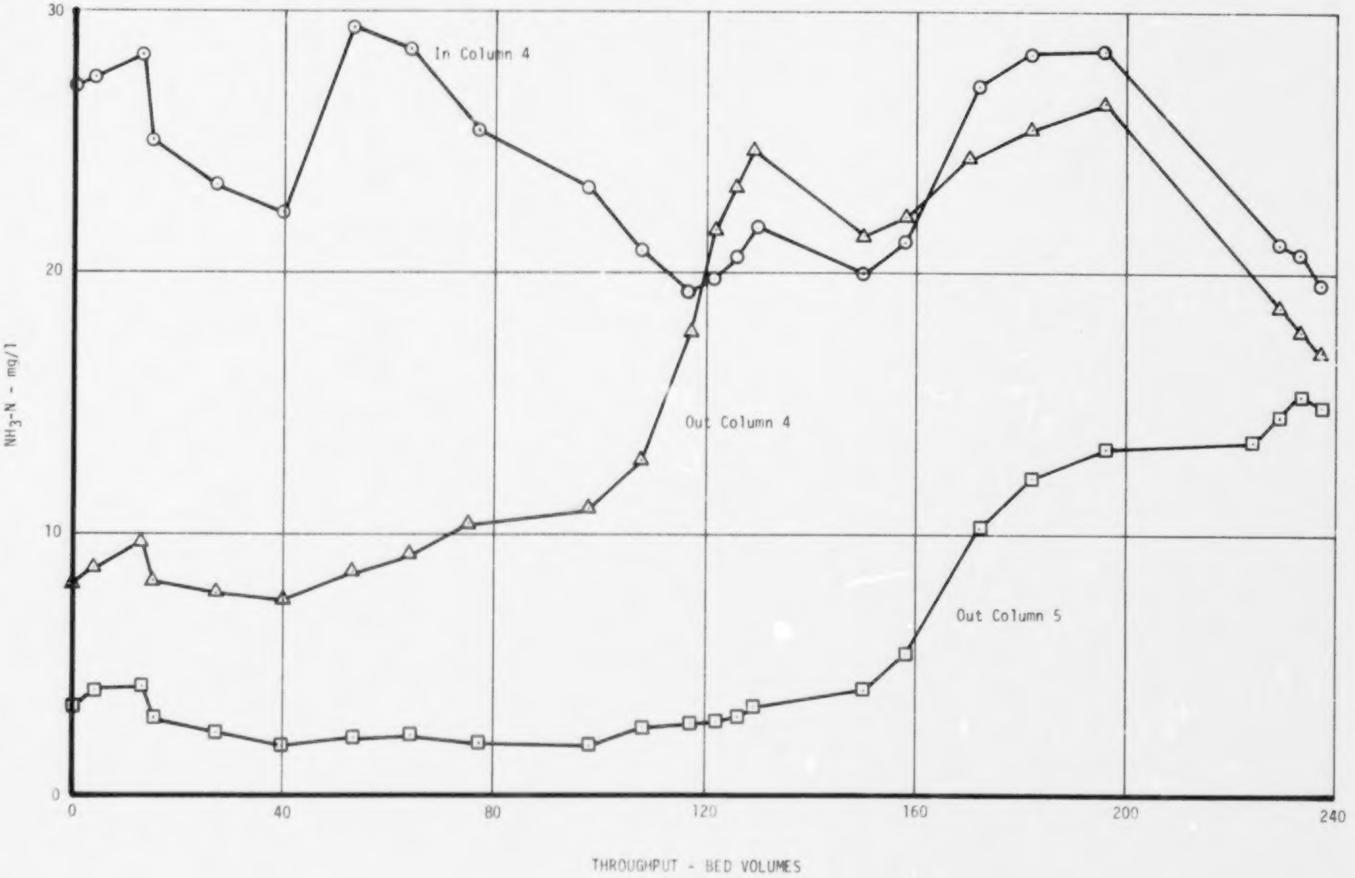


Figure 44. Ammonia breakthrough curves for zeolite columns 4 and 5 - regenerated at pH 11.

b. Scheme 2, pH 12.5 - 4.6 BV - 1 N Na

During September and October 1975 a number of full scale regeneration experiments were conducted. Ammonia elution data were collected during the regeneration of columns that had equilibrated with a process stream containing between 30 and 40 mg/l NH_3-N. These data demonstrated that regeneration at pH 12.5 (4.6 BV, 1 N Na) was much more efficient than at pH 11. Columns 5 and 6 were in service on alternate days for a total of 28 days. They were regenerated with 4.6 BV at a nominal pH of 12.5 and 1 N Na. However, prior to going into service both columns were subjected to multiple regenerations and the elution data indicated both columns were completely regenerated. For the first six days of service the column effluents were sampled on a semi hourly basis. The data for the first four days are presented in Figure 45. (Heavy precipitation on days 4 and 5 diluted the raw wastewater on days 5 and 6 and the NH_3-N concentrations were much lower than normal.) With the columns completely regenerated the initial leakage was much lower than previously observed and breakthrough did not occur until approximately 100 BV passed through the column. After one standard regeneration cycle at pH 12.5 the leakage increased and breakthrough occurred sooner although the average influent concentration did not change substantially.

Column performance continued to degrade somewhat with successive regenerations. After the fourth cycle however it appeared as if the changes in performance were more related to the influent NH_3-N concentration and flow. For the 28-day period the column influent averaged 38 mg/l NH_3-N and the effluent 12 mg/l NH_3-N. An average of 47 lbs NH_3-N was removed by the zeolite during each service cycle yielding an average dynamic exchange capacity of 0.24 me/gm.

An average of 800 lbs NaCl and 138 lbs Na_2CO_3 were added to the brine during each regeneration and brine recovery cycle. The weight of NaOH added for pH control was not quantified because the sight glass was not yet installed on the caustic storage tank. The average brine conditions are summarized in TABLE 49.

TABLE 49. AVERAGE BRINE CONDITIONS - SCHEME 2

		mg/l		
Brine	pH	Sodium	Calcium - $CaCO_3$	NH_3-N
Reclaimed	12.35	0.92 N	-	-
Waste	10.3	0.77 N	1,401	520
Desludged	12.1	-	54	-
Stripped	12.1	-	-	68

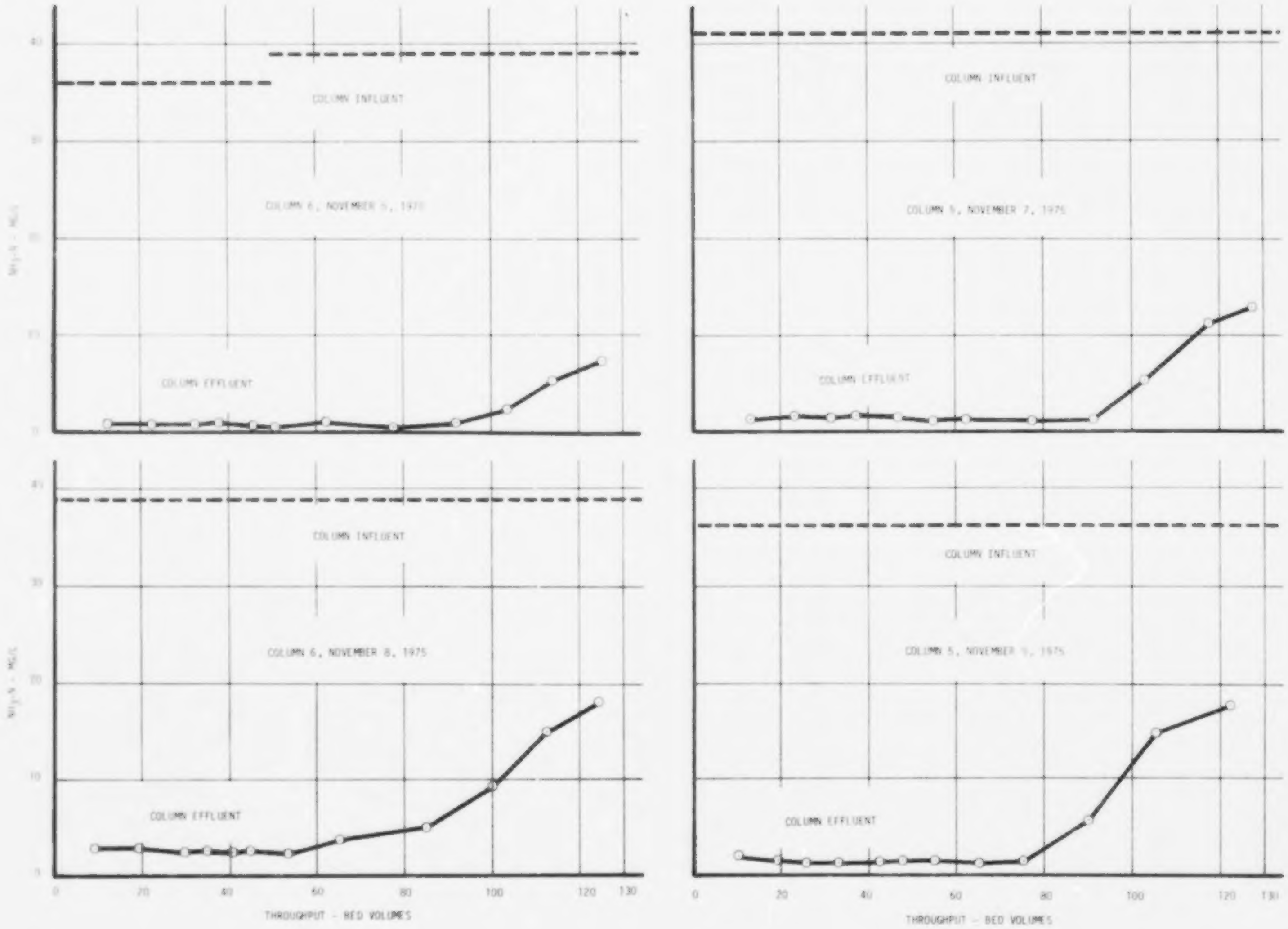


Figure 45. Ammonia breakthrough curves for columns 5 and 6 - regenerated at pH 12.5.

The average sodium concentration of the reclaimed brine was slightly less than the target value of 1.0 N and the average of the pH values was less than the goal of 12.5.

Both the reclaimed and waste brine tanks were constructed with hopper bottoms to facilitate the collection and storage of sludge which was primarily $Mg(OH)_2$ and $CaCO_3$. The suction lines of the brine pumps were located off the floor of the tanks and as a result, a minimum of approximately 1700 gal of brine and sludge remained in the tanks at all times. Based on measured depths the average brine volumes were as follows: reclaimed - 12,255 gal, waste - 11,985 gal, desludged - 11,332 gal, and stripped - 11,392 gal. These volumes and the data of TABLE 49 indicate that approximately 1,225 gm eq wt of calcium were exchanged during each complete service/regeneration cycle along with 1,524 gm eq wt NH_3-N (48 lbs). An average of 7,745 mg eq wt of sodium was lost from the brine during each column regeneration. Thus only 20% of the sodium used was exchanged for NH_3-N as follows

$$ZNH_4^+ + Na^+ \rightleftharpoons ZNa^+ + NH_4^+$$

Another 16% of the sodium is accounted for by exchange for calcium. Although no other cations were determined on a regular basis it appears that somewhat over 50% of the sodium lost between the reclaimed brine and waste brine states was wasted due to housekeeping procedures. A significant portion of this loss was most likely related to the operation of the J_1 and J_2 valves illustrated on Figure 9. The conductivity bridge originally supplied by the contractor did not have the proper range and as a result the valves were activated manually based on the volume of brine pumped. It is possible that the operators did not always activate the valves at the proper time.

c. Scheme 3, pH 13 - 4.6 BV - 1 N Na

For 13 days in December, 1975 columns 5 and 6 were in service on alternate days and regenerated at pH 13. During this period the column influent averaged 35 mg/l NH_3-N and the effluent 6.3 mg/l NH_3-N. An average of 54 lbs NH_3-N was removed each day yielding a working exchange capacity of 0.28 me/gm. The average brine conditions are summarized in TABLE 50.

TABLE 50. AVERAGE BRINE CONDITIONS - SCHEME 3

Brine	pH	Sodium	mg/l Calcium - $CaCO_3$	mg/l NH_3-N
Reclaimed	13.0	1.00 N	-	-
Waste	12.4	0.85 N	1,186	888
Desludge	12.3	-	83	-
Stripped	12.4	-	-	263

The average chemical additions per regeneration cycle to maintain these conditions were as follows: 755 lbs NaCl, 122 lbs Na_2CO_3 and 508 lb NaOH. Approximately 630 lbs sodium were added in the three forms. The average brine volumes were as follows: reclaimed - 12,232 gal, waste - 11,962 gal, desludged - 11,152 gal, and stripped - 11,130 gal.

The relatively high concentration of NH_3-N in the stripped brine was caused by a leaking pressure relief valve that allowed a significant portion of the brine to bypass the stripping column.

The removal of Ca^{++} averaged 2,510 gm eq wt per cycle and the NH_3-N removal was 1,751 gm eq wt. The sodium addition averaged 12,500 gm eq wt; however, approximately 2,610 gm eq wt were lost in the desludging operation. Approximately 45% of the sodium added to the system was unaccounted for.

d. Scheme 4, pH 13 - 4.6 BV - 0.5 N Na

The system was operated for a period of 32 days with two columns in series with one regenerated every 24 hours at pH 13 with 0.5 N NaCl. The brine strength was decreased in an attempt to reduce sodium losses. Koon and Kaufman (33) reported that increasing sodium concentrations above 0.2 N had no effect on regeneration performance. The average brine conditions are presented in TABLE 51. The average of the reclaimed brine pH value was slightly higher than the target value of 13.0.

TABLE 51. AVERAGE BRINE CONDITIONS - SCHEME 4

Brine	pH	Sodium	mg/l Calcium - $CaCO_3$	NH_3-N
Reclaimed	13.2	0.54 N	-	-
Waste	12.8	0.38 N	327	735
Desludged	12.8	-	20	-
Stripped	12.7	-	-	63

During this period the average NH_3-N concentrations for the column influent and effluent were 32 and 7.1 mg/l respectively. An average of 52 lbs NH_3-N was removed each day for an average exchange capacity of 0.27 me/gm. However, because of problems encountered with the main zeolite column bypass valve a significant portion of the flow bypassed the columns and it is possible that the calculated exchange capacity would have increased if all of the flow had passed through the columns.

The average chemical requirements for brine makeup were as follows: 300 lb NaCl, 30 lb Na_2CO_3, and 500 lb NaOH. A total of approximately 420 lb sodium was added per regeneration cycle compared to 630 lb per cycle when 1.0 N NaCl brine was used.

The average volumes for the brine states were as follows: reclaimed - 12,230 gal, waste - 11,350 gal, desludged - 10,670 gal, and stripped - 10,860 gal. Based on these values and the data in TABLE 51 approximately 681 gm eq wt calcium were removed during each cycle. The NH_3-N removal averaged 1,686 gm eq wt. The measured sodium addition amounted to approximately 8,300 gm eq wt; however, approximately 980 gm eq wt were lost during desludging. Thus approximately 60% of the sodium was unaccounted for.

The data indicate that the zeolite capacity for NH_3-N was not significantly affected by the reduction of sodium concentration; however, there was a significant reduction in the observed calcium removed based on the brine analyses.

Attrition losses of clinoptilolite are reported to increase as the regeneration pH is increased and reach approximately 0.5% weight loss per regeneration cycle at pH 12.5 (33). Prior to regenerating the zeolite beds at pH 13, columns 4, 5, and 6 were drained and the zeolite surface located. The same measurements were made after a significant number of regenerations at pH 13 were made. The results are presented in TABLE 52.

On a volume basis it did not appear that significant losses had occurred. It is possible however, that some weight loss did take place.

TABLE 52. ZEOLITE LOSSES FOR pH 13 REGENERATIONS

Column No.	Number of Regenerations at pH 13	Zeolite Loss-in
4	7	- 4
5	14	2.5
6	14	- 0.5

When the ammonia removal system was first put into service in May 1975 a measurable pH increase was observed across the zeolite columns. When the regeneration pH was increased to 13, the pH increase was dramatic. The effluent composite pH was a high as 9.0 and the instantaneous effluent pH was as high as 12.0. The column effluent pH would reach this high value immediately after a freshly regenerated column went on line and then gradually decrease. For the period December 10, 1975 to March 24, 1976 the average of the effluent pH values (obtained from flow composite samples) was 8.2.

On March 24 an acid neutralization system was put into service to preclude pH violations (NPDES limit - 8.5) and to eliminate the return of rinse water to the treatment processes. After a column was regenerated and refilled the rinse cycle was started. Water was pumped from filtered water storage tank number 2 through the column, downflow, at a rate of 150 gal/min and discharged in the train 1 side of filtered water storage tank 1 where acid was added to reduce the pH to approximately 7.5. The pH control equipment on clarifier No. 1 was modified to achieve this task. The neutralized flow was then pumped through train 1 piping, bypassing the treatment units, and discharged to the train 1 side of filtered water storage tank No. 2. From this point the flow was pumped directly to the outfall line. It was found that rinse times on the order of 360 minutes were required to bring the pH of the column effluent to a value below 8.5.

The above rinse procedure was terminated on July 3, 1976 when the pH controllers serving both clarifiers were modified by the contractor to utilize the clarifier flow signals as well as pH inputs. This modification was an attempt to improve pH control in the clarifier (see Figure 29). For reasons not yet known the contractor and instrument supplier were not capable of successfully executing the modifications; however, no further attempt was made to acid neutralize the zeolite rinse waters.

e. Scheme 5, pH 12 - 4.6 BV - 0.5 N Na

The pH 13 regenerations were terminated because it was difficult to maintain the effluent pH below the required 8.5 and the cost of the sodium hydroxide used for pH control made the high pH regenerations impractical. For a 14-day period the regeneration pH was lowered to 12 but no other changes were made in the operating procedures. During this period the NH_3-N concentrations for the zeolite column influent and effluent averaged 26 mg/l and 11 mg/l respectively. The NH_3-N concentration of the plant effluent, however, was substantially higher than the column effluent for 5 of the 14 days and averaged 13 mg/l. It was obvious that several valves were not seating properly at all times and, as a result, a significant portion of the flow bypassed the columns. On several occasions both the chloride concentration and conductivity of the plant effluent increased dramatically during the zeolite regeneration process. This indicated that brine was leaking to the effluent through supposedly closed valves.

Because of the problems encountered with the valves it is not possible to interpret the performance data. At that time, however, it was not possible to repair or replace the defective valves. The elevation of a considerable portion of the plant outfall is higher than the plant floor and that portion contains approximately 450,000 gal. At that time there was no positive shut off valve to isolate the plant from the outfall and it was not known if the installed check valves would seal properly. The outfall was drained and a positive shut off valve was installed June 29, 1976.

f. Scheme 6, pH 12 - 8.2 BV - 0.5 N Na

The brine transfer piping was changed to allow direct transfer from the waste to the reclaimed tank. This modification facilitated regenerations with more than the previously fixed 4.6 BV of brine. The columns were operated for a period of 84 days during which time the average regenerant volume was 8.2 BV. The data obtained during the first 18 operating days are suspect because of the problem associated with the leaking valves; however, at least one potential problem was identified. During these 16 days of operation the columns were rinsed at a rate of 200 gpm (4 gpm/sq ft) in the upflow mode for approximately 2 hrs. The rinse water was discharged to the backwash pit and recycled to the headworks. On July 22, 1976 a substantial amount of material that appeared to be zeolite was found in the backwash pit. On the following day columns No. 4, 5, and 6, which had been used during that period, were drained and opened. Depth measurements indicated that a total of approximately 115 cu ft of zeolite was lost. Subsequent upflow rinsing was accomplished at rates$\leq$100 gpm (2 gpm/sq ft).

The valve repair program was completed in mid October 1976. The average column influent and effluent NH_3-N concentrations for the remaining 68 days where 44 mg/l and 12 mg/l respectively. Removals ranged from 31 to 89 lbs NH_3-N per day and averaged 60.4 lb/day yielding an average exchange capacity of 0.31 me/gm. The average brine conditions are summarized in TABLE 53 and the average chemical additions per regeneration were as follows: 480 lbs NaCl, 137 lbs Na_2CO_3, and 127 lbs NaOH. A total of approximately 320 lbs sodium was added per regeneration cycle.

The average volumes recorded for the brine conditions were as follows: reclaimed - 12,320 gal, waste - 12,090 gal, desludged - 11,610 gal and stripped - 11,640 gal. Based on these volumes approximately 2,900 gm eq wt calcium were removed during each cycle and 710 gm eq wt sodium were lost by desludging. The sodium addition averaged 6,300 gm eq wt per cycle and the NH_3-N removal averaged 1,960 gm eq wt. Thus approximately 88% of the sodium addition was accounted for.

TABLE 53. AVERAGE BRINE CONDITIONS - SCHEME 6

Brine	pH	Sodium	mg/l Calcium - $CaCO_3$	mg/l NH_3-N
Reclaimed	12.0	0.48 N	-	-
Waste	10.0	0.39 N	1,357	544
Desludged	11.8	-	76	-
Stripped	11.8	-	-	90

It is reported that for cations normally encountered in domestic wastewaters the preference of clinoptilolite for ammonium ion is exceeded only by its preference for potassium ion. (36) If the column influent contains significant levels of potassium the capacity for ammonium ion may be reduced substantially. In such a situation the potassium level in the reclaimed brine may build up to such a level as to preclude complete sodium regeneration.

During March and April 1977 the potassium concentrations of the process and brine samples were determined. Prior to initiating the potassium analyses the brine storage tanks were drained and cleaned and fresh brine made up. The potassium concentration of the reclaimed brine increased from approximately

10 to 250 mg/l after eight regeneration cycles. After that time the concentration varied between 100 and 250 mg/l. This variation was most likely related to the volume of makeup water required on any particular day. For a 20-day period after the potassium level increased the concentration varied from 85 to 250 mg/l and averaged 169 mg/l.

During this same period the potassium concentration for the INAEC, AEC_1 and AEC samples averaged 14.5 mg/l, 19.5 mg/l and 13 mg/l respectively. The net potassium removal averaged 3.6 lb/day or 42 gm eq wt/day which is equivalent to approximately 2% of the NH_3-N removed. It does not appear that the nitrogen removals were substantially reduced because of the net potassium removals.

g. Scheme 7, pH 12 - 14 BV - 0.5 N Na

The regeneration procedure was modified to include a 3rd and 4th pass of the waste brine after pH adjustment - see Section IV. G.3. The columns were operated for a total of 35 days using this procedure. The sampling system was modified to provide a composite sample of the primary column effluent. The NH_3-N concentrations of the primary column influent and effluent and secondary column effluent averaged 35 mg/l, 23 mg/l and 4.6 mg/l respectively. The average flow treated was 123 BV (based on one column) and the average removal was 71 lbs/day. The zeolite capacity for these conditions averaged 0.37 me/gm.

During this period the average regenerant volume was 14.0 BV. The brine characteristics are summarized in TABLE 54 and the average volumes for the brine conditions were as follows: reclaimed - 12,234 gal, waste - 12,276, desludged - 11,804 gal and stripped - 11,640 gal. The chemical additions averaged 450 lb NaCl, 112 lb Na_2CO_3 and 310 lb NaOH per regeneration. Approximately 68% of the total sodium addition of 400 lbs was accounted for in terms of ion exchange for NH_4^+ and Ca^{++} and loss during the desludging operation.

TABLE 54. AVERAGE BRINE CONDITIONS - SCHEME 7

			mg/l	
Brine	pH	Sodium	Calcium - $CaCO_3$	NH_3-N
Reclaimed	12.0	0.49 N	-	-
Waste	10.9	0.39 N	1,077	730
Desludged	-	-	49	-
Stripped	-	-	-	116

h. Scheme 8

The ammonia stripping procedure was modified slightly during 27 days of operation; however, during this period the regeneration and operating procedures for the zeolite columns were as described in the previous section. The metered steam application rate for all previous stripping operations was maintained at 1 lb steam per gal brine. During this period the steam rate was reduced to 0.75 lb/gal. As expected the NH_3-N concentration of the stripped brine increased and averaged approximately 270 mg/l. It was not anticipated that the increased NH_3-N level would have a measurable effect on the regeneration efficiency because the brine pH was maintained at 12.0 for all four passes.

The NH_3-N concentrations averaged 27 mg/l, 15 mg/l, and 3 mg/l for the primary column influent and effluent and the secondary column effluent respectively. The flow averaged approximately 112 BV (based on one column). The calculated zeolite capacity for these conditions is 0.27 me/gm. Unfortunately both the plant flow and NH_3-N concentration were significantly lower than during the previous period and the data are not easily interpreted.

Based on an isotherm developed in the laboratory the exchange capacity at equilibrium conditions is decreased by approximately 13% when the NH_3-N concentration is decreased from 35 to 27 mg/l. In addition, at the lower flow rates the columns were most likely furtherfrom a state of complete exhaustion than during the previous period (Scheme 7). Based on these results it does not appear that reducing the steam rate by 25% produced a significant effect.

i. Scheme 9

One additional modification was made to the NH_3-N stripping procedure. For a 21-day period during which the zeolite columns were operated as in the previous section the NH_3-N was stripped from the waste brine on alternate days. The NH_3-N concentration of the stripped brine averaged 420 mg/l. The NH_3-N concentration in the reclaimed brine ranged from 350 to 1,070 mg/l and averaged 700 mg/l. The pH of the reclaimed brine was again maintained at 12.0 and the ammonia was thus quite volatile. The dissociation constant for aqueous ammonia solutions is approximately 1.7×10^{-5} at room temperature. (35) At pH 12 at equilibrium conditions no measurable portion of the NH_3-N is in the form of ammonium ion (NH_4^+) to interfere with the ion exchange process.

During this period the NH_3-N concentrations of the primary column influent and effluent and the secondary column effluent averaged 30 mg/l, 23 mg/l, and 6.6 mg/l respectively. The treated flow averaged 121 BV and the calculated exchange capacity was 0.27 me/gm. When these values are compared to the results obtained during operating schemes 7 and 8 it appears that the zeolite performance was measurably degraded by the high residual NH_3-N in the regnerant. Severe odor problems were also associated with this modification because of the high concentration of volatile NH_3 formed in the reclaimed brine.

3. Summary of Performance

The data collected during 263 days of operation of the ammonia removal system are summarized in TABLE 55. For the eight operating schemes considered, the NH_3-N removal efficiency varied from 41% to 89% and averaged approximately 72% while treating approximately 122 BV flow. The exchange capacity and removal efficiency were significantly affected by both the regenerant volume and pH. It did not appear that the sodium content of the brine had a significant effect when the concentration was maintained above 0.5N.

The chemical requirements for brine makeup and conditioning varied substantially for several of the regeneration schemes. Chemical consumption was reduced substantially by reducing the sodium strength from 1.0N to 0.5N and by conducting multiple pass regenerations at pH 12.0.

Although significant zeolite losses were experienced as a result of prolonged upflow rinsing at a rate of 4 gpm/sq ft, no zeolite attrition was directly attributable to high pH regenerations. However, when the regenerations were conducted at pH values above 11 a significant volume of water was required to rinse the zeolite to ensure that the pH of the column effluent was less than 8.5 which was the plant's upper limit according to the discharge permit. Notwithstanding the permit conditions, the rinse was required to ensure that the major portion of the ammonia was in the form of NH_4^+ and thus available for exchange and removal from the wastewater.

4. Cost

The cost data summarized in TABLE 56 include the charges against the four cost centers (070, 120, 130 and 140) associated with the ammonia removal system (see TABLE 12). For the 226 days of operation that fell within the grant period the total operating cost was approximately $1,425 per mil gal.

Approximately 74% of the costs were related to operating labor plus maintenance labor and materials. The operating labor requirement was high because of the poor reliability of several of the installed automated control loops and because the modifications to the regeneration procedure required manual control. The maintenance costs are associated primarily with: piping modifications, valve repair and replacement and repair and cleaning of the heat exchangers. The major portion of these costs would not have been incurred if the facility design was based on information now available. The operating costs reported are thus not representative of a well design facility treating the same flow.

The chemical costs represent 14.5% of the recorded operating cost. Both the sodium chloride and soda ash were purchased by the bag because bulk storage facilities were not provided at the plant. The average delivered unit prices for salt and soda ash were $0.0215/lb and $0.069/lb respectively. The unit cost for caustic was also relatively high, $0.128/lb, because the small storage tank required delivery of partial loads by tank truck. In addition the total power cost for the effluent pumps was charged against the ammonia removal process and further inflated the cost.

TABLE 55. AMMONIA REMOVAL SYSTEM PERFORMANCE SUMMARY

Scheme	Regenerant			NH_3-N, mg/l		Flow	Average NH_3-N Exchange Capacity	lbs per Regeneration Cycle			
	pH	BV	Na-N	In	Out	BV/day	me/gm	NaCl	Na_2CO_3	NaOH	Σ Na
1	11	4.6	1	29	17	98	0.12	-	-	-	-
2	12.5	4.6	1	38	12	98	0.24	800	138	-	-
3	13	4.6	1	35	6.3	102	0.28	755	122	508	600
4	13	4.6	0.5	32	7.1	104*	0.27	300	30	500	401
6	12	8.2	0.5	33	12	150	0.31	480	137	127	300
7	12	14	0.5	35	4.6	123	0.37	450	112	310	380
8**	12	14	0.5	27	3.0	112	0.27	422	104	320	373
9***	12	14	0.5	30	6.6	121	0.27	400	101	254	327

* unknown quantity of flow bypassed zeolite columns

** steam rate reduced to 0.75 lb/gal brine

*** strip on alternate days with reduced steam rate

TABLE 56. AMMONIA REMOVAL COSTS

Category		Quantity	$	%
Direct labor -	hr	4,231	28,787	30.6
Maintenance labor -	hr	1,853	15,804	16.8
Benefits		-	8,917	9.5
Acid -	lb	-	280	0.3
Sodium chloride -	lb	163,350	4,327	4.6
Soda ash -	lb	25,690	1,763	1.9
Sodium hydroxide -	lb	54,196	6,937	7.4
Electricity -	KWH	209,170	4,603	4.9
Gas -	MCF	5,315	6,263	6.6
Materials & supplies		-	16,335	17.4
Total		-	94,016	100.0

E. CARBON REGENERATION

1. Efficiency

Successful carbon regenerations were made as early as October 1974 as illustrated by the data summary in TABLE 57. Although the temperature control system functioned in an erratic manner and difficulty was encountered in maintaining the proper air to gas ratio in the burners, the system was operated and manually adjusted to regenerate the spend carbon. The values for apparent density and iodine number of the regenerated carbon indicate that 2 of the 22 regenerations can be rated poor. The regeneration completed December 4, 1974 was conducted without the aid of the apparent density apparatus which was delivered to the plant the first week in January 1975. During the first 5 regenerations the operators had no means to evaluate the process during operation. The second poor regeneration, ending March 20, 1975, was caused by several failures of the burners, induced draft fan and scrubber.

Prior to August 1975 when the variable speed drive was installed on the screw conveyor (Figure 8) the operator had little control of the regeneration process except for varying the steam rate and temperature set points. The temperature control system did not function properly until it was rewired

TABLE 57. ACTIVATED CARBON REGENERATION DATA SUMMARY

Date Regeneration Completed	Carbon Feed Rate cu ft/hr	Apparent Density		I_2 Number		Ash - %		Volatile		Average Hearth Temp - °F			
		Spent	Regen.	Spent	Regen.	Spent	Regen.	Spent	Regen	1	2	3	4
Oct. 25, 1974	-	-	-	-	942	-	-	-	-	-	-	-	-
Nov. 13, 1974	-	-	-	576	833	-	-	-	-	-	-	-	-
Nov. 15, 1974	-	-	-	558	825	-	-	-	-	-	-	-	-
Dec. 4, 1974	-	0.60	0.55	538	786	-	-	-	-	-	-	-	-
Dec. 7, 1974	-	0.55	0.50	679	919	-	-	-	-	-	-	-	-
Jan. 15, 1975	-	0.54	0.46	732	990	-	-	-	-	-	-	-	-
Jan. 18, 1975	-	0.54	0.49	750	906	-	-	-	-	-	-	-	-
Jan. 22, 1975	-	0.56	0.49	656	932	-	-	-	-	-	-	-	-
Jan. 26, 1975	-	0.53	0.49	716	924	-	-	-	-	-	-	-	-
March 20, 1975	10.8	0.54	0.52	721	796	-	-	-	-	909	1466	1525	1639
April 27, 1975	8.4	0.49	0.46	870	1026	-	-	-	-	891	1430	1460	1593
May 2, 1975	8.5	0.51	0.47	868	924	-	-	-	-	850	1301	1400	1564
May 5, 1975	-	0.53	0.47	777	959	-	-	-	-	922	1428	1472	1602
Sept. 20, 1975	11.8	0.51	0.48	771	910	-	-	-	-	947	1428	1539	1589
Jan. 12, 1976	7.5	0.58	0.50	528	883	3.2	4.3	14.1	1.9	1030	1628	1561	1618
Jan. 15, 1976	8.3	0.57	0.50	658	903	4.2	4.6	9.0	2.0	1025	1590	1606	1639
Jan. 20, 1976	-	0.56	0.52	680	855	4.8	4.3	3.7	1.3	1048	1647	1649	1684
Sept. 17, 1976	10.0	0.54	0.48	700	1049	2.8	3.2	6.6	2.0	916	1512	1564	1623
Nov. 26, 1976	7.2	0.58	0.51	583	874	4.2	5.1	10.1	2.2	1005	1555	1563	1638
Dec. 4, 1976	4.5	0.62	0.51	491	944	5.5	7.1	13.5	2.4	1115	1645	1673	1729
Dec. 9, 1976	7.6	0.60	0.50	456	772	4.3	5.0	15.0	2.8	993	1546	1552	1598
Dec. 23, 1976	11.1	0.58	0.50	512	807	4.8	4.8	9.6	2.9	980	1566	1600	1574
Original Carbon inventory	-	-	0.48	-	980	-	-	-	-	-	-	-	-
Makeup carbon	-	-	0.49	-	1166	-	6.3	-	1.7	-	-	-	-

in January 1976. Prior to that time large temperature excursions were commonplace. For the regeneration ending May 2, 1975 the standard deviations of the 80 hourly temperature readings were 143°F, 26°F, 209°F, and 196°F for hearths 1, 2, 3, and 4 respectively. The standard deviations fell in the range of 20°F to 60°F after the wiring modifications were completed.

Because of the difficulty in identifying carbon batches it was not possible to make this type of determination. Although it is possible to trace the movement of carbon batches from column to column through each transfer and regeneration sequence their integrity was not maintained. Two of the initial carbon batches were eliminated when they were used to top off other columns and significant quantities of virgin carbon were also used to top off columns. When the replacement carbon is taken into consideration the carbon in inventory as of December 23, 1976 had been regenerated an average of only 3 times. The COD removal data previously presented do not indicate that carbon age was significant.

The iodine number is an indication of the capability of an activated carbon to adsorb low molecular weight substances.(2) Several investigations indicate that although the iodine number of the carbon decreases with successive regenerations its capacity to adsorb wastewater organics may increase. (25)(37) Experience at Rosemount indicates that the iodine number of the regenerated carbon is more closely related to the state of the spent carbon and the regeneration conditions than the number of previous regenerations.

2. Losses

When the total carbon loss is divided by the number of regenerations the average loss is approximately 11.4% per regeneration cycle. As previously stated however, it is estimated that at least 50% of the loss was associated with the routine carbon backwash procedures. During the initial regenerations, significant quantities of carbon were discharged to the backwash pit because of a design deficiency. The spent carbon storage tank discharged through a slide gate to a small hopper (volume approximately 1 cu ft) which fed the screw conveyer. The hopper was partially flooded to maintain the furnace seal. The water level was controlled by a short section of weir. The dewatered carbon did not discharge from the storage tank by gravity because of interparticle bridging. Water was added in the lower section of the storage tank to maintain a carbon discharge; however, the discharge was not steady. As a result the small hopper was overloaded and carbon overflowed the weir and discharged to the backwash pit. Although the drain was covered with a high-hat screen the losses were significant until the operators developed a feel for balancing the discharge from the storage hopper and the screw transport rate.

Because the carbon furnace was used infrequently and was subjected to a corrosive atmosphere of both H_2S and moist activated carbon the startup procedure involved repair or recalibration of the gas and air control devices. During one of the initial regenerations the carbon bed caught fire and the hearth temperature went wild. After the furnace was shut down and the charge rabbled out it was determined that several of the gas/air metering devices had malfunctioned and allowed considerable excess oxygen in the furnace. After this experience the startup procedure included the determination of the oxygen content on the hearths prior to initiating carbon feed and adjusting the metering devices when the oxygen content increased beyond 2% by volume. Because of the location of the metering devices the calibration and repair procedures were extremely time consuming. Fortunately it was determined that firing the furnace once each month eleiminated the need for the recalibration procedure and as a result the oxygen levels were no longer determined. It is possible that a measurable quantity of carbon was burned in the furnace since no direct column measurements were made before and after regeneration.

During recent regenerations (September 1977) significant quantities of carbon were found to accumulate in the backwash pit toward the end of the regeneration. The regenerated carbon is hydraulically transported to an off-line storage column with an eductor. The carbon is deposited in the column and water is removed through the underdrain system. This column is also equipped with an unscreened overflow near the top. It appears that as the column fills with carbon the water level rises to the overflow level and carbon which is discharged to the top of the column floats out the overflow. When this problem is rectified the carbon losses will be minimized.

Although no direct measurements were made it was not felt that significant losses were related to mechanical attrition. The regenerated carbon when backwashed did not appear to contain a significant weight of carbon fines. An attrition loss of 0.3% per cycle is reported by Von Dreusche for movement through an eight hearth furnace.(39)

3. Cost

A total of approximately 140,000 lbs carbon were reactivated in the 9 regenerations that took place during the demonstration period. The total recorded cost was approximately $21,000 as summarized in TABLE 58. The costs associated with the evaluation of the regeneration in terms of apparent densities, iodine numbers, etc. and the cost of make up carbon were not included in this cost center.

TABLE 58. CARBON REGENERATION COSTS

Category	Quantity	$
Direct labor - mhr	1,311	9,684
Maintenance labor - mhr	565	4,730
Benefits	-	2,722
Gas - MCF	1,286	2,060
Materials and supplies	-	1,486
Contract maintenance	-	298
Total	-	20,980

XII. COSTS

A. OBSERVED - .25 mgd

The total cost for operating the Rosemount AWTP during the two-year demonstration project was approximately $1,005,000. Of this total approximately $100,000 was directly attributable to the evaluation of performance. The actual operating and maintenance costs are summarized in TABLE 59. The costs associated with process modifications are included although they might not be considered normal O & M items. Although sludge hauling costs are included no charges were allocated for sludge disposal.

TABLE 59. OPERATING AND MAINTENANCE COST SUMMARY

Category	$	% total cost
Labor	587,721	64.9
Chemicals	60,452	6.7
Materials & supplies	80,296	8.9
Energy & fuel	70,092	7.7
Contract maintenance	64,022	7.1
Sludge hauling	32,325	3.6
Miscellaneous	10,103	1.1
Total	905,011	100.0

The total raw wastewater flow during the two-year period was approximately 177 mil gal and the estimated cost for treatment is $5.11 per 1,000 gal. This unit cost is extremely high because the plant capacity and flow are low enough to preclude any economy of scale and yet the plant complexity required full staffing. Assume, for the sake of simplicity, that the costs for chemicals, sludge hauling and energy and fuel are related to treated flow on a 1:1 basis. If the plant had operated at its design capacity, 0.6 mgd, for the two-year period the average unit cost of treatment would decrease by approximately 49% to $2.62 per 1,000 gal.

Salaries, wages, fringe benefits and payment for vacation and sick leave amounted to approximately 65% of the total O & M cost. The recorded distribution of hours amoung the 17 cost centers is presented in TABLE 60. Cost centers 10 through 140, which are related to actual treatment units, accounted for approximately 42% of the direct and 65% of the maintenance labor times.

More direct labor was expended on clarification than any of the other treatment processes; however, the cost centers associated with carbon regeneration and the ammonia removal systems were much more labor intensive. The carbon regeneration system, cost center 110, was operated for 103 days including startup times and required approximately 12.7 mhr per day. The entire ammonia removal system, cost centers 70, 120, 130, and 140 was in operation for 226 days during the grant period and averaged 18.7 mhr per day. The maintenance labor requirements for the carbon regeneration system and the entire ammonia removal system were significant and averaged 5.5 mhr and 8.2 mhr respectively.

B. PREDICTED - 10 mgd

Estimates for the operating and maintenance costs for a 10 mgd facility are presented in the following paragraphs. It is assumed that this new independent physical chemical treatment facility would include the design modifications discussed in the following sections and would thus be somewhat more efficient than the Rosemount facility. The following assumptions were used to construct the cost estimates which were based on December 1977 costs at Minneapolis-St. Paul.

Effluent quality	BOD - 10, SS - 5, Total P - 1, NH_3-N - 3.
Clarification	lime to pH 10.5, 20 mg/l $FeCl_3$, neutralize to pH 7.5
Carbon Adsorption	remove 25 mg/l soluble COD, capacity = 0.4 lb COD/lb carbon, add minimum of 5 mg/l NO_3-N, carbon losses limited to 5% per regeneration cycle
Ion exchange	remove 15 mg/l N, zeolite capacity = 0.2 me/gm regenerate with 20 BV brine (a) reclaim brine by steam stripping (b) reclaim brine by biological nitrification as described by Semmens (39) with return of a portion of nitrified brine to carbon columns to satisfy NO_3 requirement for sulfide control.

TABLE 60. LABOR REQUIREMENT FOR COST CENTERS

Cost Center No.	Cost Center Title	Labor - mhr * Direct	Labor - mhr * Maintenance
10	Preliminary Treatment	1,666	149
20	Influent Pumping Station	963	284
30	Clarification & Phosphorus Removal	5,816	936
40	Filtration 1	1,486	340
50	Carbon Adsorption	1,957	120
60	Filtration 2	1,184	346
70	Ammonia Removal	160	168
80	Chlorination	454	88
90	Effluent Disposal	178	203
100	Sludge Handling & Disposal	327	341
110	Carbon Regeneration	1,311	565
120	Clinoptilolite Regeneration	2,679	1,609
130	Ammonia Recovery System	1,102	38
140	Ammonia Disposal	290	50
150	Laboratory Services	6,310	492
160	Buildings & Grounds	11,945	1,947
170	Indirect Services	9,087	396

* does not include vacation or leave time

The labor requirements presented in TABLE 60 should not be considered to be representative of the requirements for a well designed and constructed treatment facility. For the 10 mgd facility it is estimated that 21 full time positions would be required as summarized in TABLE 61.

TABLE 61. ESTIMATED FULL TIME STAFF FOR 10 MGD FACILITY

Position	Number	Shift
Operator	4	all
Assistant operator	4	all
Operating attendant	4	all
Electrician	2	day
Pipefitter	2	day
Machinist	1	day
Chemist	1	day
Laboratory technician	1	day
Secretary	1	day
Superintendent	1	day

Based on wage and salary rate schedules available December 1977 the total labor cost would be $453,300 including benefits and leave. The unit cost would be approximately $0.124 per 1,000 gal.

The observed chemical costs averaged approximately $0.34 per 1,000 gal. If the ammonia removal system had been in service 100% of the time this cost would have increased substantially. The observed chemical costs, however, should not be considered representative because purchases were made in small lots and no economy of scale was obtained. The estimated chemical costs for the 10 mgd plant are summarized in TABLE 62 for the two methods of brine recovery considered.

All of the chemical costs are based on bulk delivery rates available to the MWCC in December 1977.

It is assumed that materials and supplies, contract maintenance and miscellaneous expenses would total $200,000 per year or approximately $0.055 per 1,000 gal. The fuel and power cost estimates are dependent on the method of brine recovery. For the method of steam stripping total annual fuel and

and power costs are $989,400 and $391,400 respectively or a total of $0.378 per 1,000 gal. For biological regeneration the estimates are $78,600 and $409,000 or $0.133 per 1,000 gal.

TABLE 62. ESTIMATED CHEMICAL COSTS - 10 MGD FACILITY

Chemical	Method of Brine Recovery Steam Stripping	 Biological
Lime	424,900	424,900
Ferric chloride	45,600	45,600
Sulfuric acid	136,500	136,500
Sodium nitrate	67,200	0
Activated carbon	58,000	58,000
Chlorine	34,500	34,500
Soda ash	105,100	198,000
Sodium hydroxide	117,900	0
Total cost $	989,700	897,500
$/1000 gal	0.271	0.246

The cost summary of TABLE 63 indicates that the treatment can be accomplished in the range of approximately $0.56 to $0.83 per 1000 gal when the cost of sludge disposal is not included. Because no sludge disposal cost data were generated at the Rosemount facility no estimate is included in this report; however, it is assumed that land disposal of liquid or cake would be utilized.

TABLE 63. COST SUMMARY - 10 MGD FACILITY

Category	Method of Brine Recovery	
	Steam Stripping	Biological
Labor	453,300	453,300
Chemicals	989,700	897,500
Materials & supplies	165,000	165,000
Energy and fuel	1,380,800	487,600
Contract maintenance	25,000	25,000
Miscellaneous	10,000	10,000
Total cost - $	3,023,800	2,038,400
- $/1,000 gal	0.83	0.56

XIII. DESIGN RECOMMENDATIONS AND CONSIDERATIONS

A. GENERAL

Although the Rosemount AWTP has generally produced a high quality effluent during its four years of operation the costs associated with several of the unit processes, in terms of manhours and chemical consumption, have been relatively high. It does not appear that independent physical chemical treatment will be cost effective in the 0.2-0.6 mgd range; however, larger plants of this type may be effective both in terms of cost and pollution abatement if properly designed and constructed. The following recommendations are based on the four years operating and maintenance experience and thus are based on relatively long-term observations.

B. SAMPLING AND MONITORING

The sampling system should be included with the process design and not as an afterthought. Sample valves should be installed both upstream and downstream of every process element so that the performance of each carbon column, for example, can be evaluated and malfunctions identified without delay. All sample streams should be conducted to a centrally located laboratory where 24 hr flow composite samples can be constructed. Grab samples or continuous monitors can be used to document diurnal variations and the effects of flow and loading fluctuations. If continuous monitors are used on a routine basis they should output to a digital computer system for data storage and retrieval, rather than to strip chart recorders. Analog data stored on paper are not readily accessible because the charts are cumbersome and the data reduction process is labor intensive. If the output of monitors is not to be used on a routine basis it would be better to use a grab sampling program to obtain time related data as required.

C. CLARIFICATION

The efficiency of the chemical clarification process in terms of chemical and labor requirements as well as effluent quality should improve by using the conventional water treatment flow scheme of rapid mix, flocculation followed by sedimentation rather than a solids contact type device. If lime is used, the rapid mix chamber must provide sufficient mixing intensity and detention time to ensure complete dissolution of the sparingly soluble $Ca(OH)_2$. Mechanical mixers are not recommended as experience indicates that they foul rapidly with rags and other debris. If grit removal is to be provided the lime could be added to and mixing accomplished in an erated grit chamber/s. The ferric chloride addition point should be near the discharge to the flocculation basin.

The flocculation basins should be equipped with variable speed flocculators and a solids recycle stream from the settling basin to allow for continuous optimization as the wastewater characteristics change and/or chemical usage is modified. Solids recycle may help to stabilize the flow with respect to calcium carbonate precipitation. Calcium carbonate precipitation on the clarifier launder, which required regular cleaning at Rosemount, and other structural surfaces should be significantly reduced.

D. FILTRATION

On numerous occasions multiple air scour and water backwash procedures along with the addition of chlorine or sodium chloride were required to clean the filter media. On one occasion the surface of the media was hand raked to remove scum that had developed and was not removed by the backwashing procedure. Based on these results it is recommended that a surface wash be incorporated in the filter design and that the rates of both the air scour and water backwash be controllable along with their durations. It would also be advisable to provide for the addition of a cleaning agent to the filter bed as required. Samples of the spent backwash water should also be readily available and perhaps monitored for turbidity to control backwash volumes.

E. CARBON ADSORPTION

If the carbon column feed stream contains a significant concentration of sulfate, problems will most likely be caused by the generation of hydrogen sulfide. This problem must be addressed by controlling the H_2S generation, with NO_3 addition for example, or by oxidizing the H_2S generated. Cost analyses will dictate the method of approach for any particular facility.

The cost of replacing activated carbon can be significant and it thus appears that the design of the contractors should eliminate the possibility of losses of specification size carbon particles during the service and backwash cycles. This would require that both the inlet/s and outlet/s be properly screened. As with the filters both the rates of air scour and backwash along with duration should be readily controllable and the backwash water monitored for solids.

Because moist activated carbon corrodes mild steel rapidly the vessels, if constructed of mild steel, must be coated. Based on the experience at Rosemount the reliability of coal tar epoxy paint which has been recommended (2) is suspect.

F. CARBON REGENERATION

The carbon transport, storage, and furnace feed systems must be designed to eliminate the loss of specification size carbon particles. This can be accomplished by properly screening all drains and overflows.

Combustion losses in the furnace can be mimized by careful control of excess oxygen. This can perhaps be best accomplished by direct metering of both the gas and air streams rather than an aspirator meter-mix device used on the furnace at Rosemount. If the furnace is not operated on a continuous basis the start up procedure should include a determination of oxygen levels on all hearths at operating temperature prior to carbon feed. This requires sampling ports and an oxygen analyzer that is reliable in the 0 to 2% range.

The combustion air intake should be removed from the carbon handling area and filtered to preclude the entry of carbon fines to the metering system which may in turn cause corrosion problems. If hydrogen sulfide is generated in the plant building it may be advisable to draw combustion air from outside the building, even if it must be preheated in the winter months. Corrosion problems were encountered with the fuel and air metering valves at Rosemount prior to moving the air intake from the floor near the furnace to near the ceiling of the building.

All of the carbon vessels should be equipped with a device that can be used to measure carbon depth and thus allow routine inventory determinations. The devices may vary from a plate and staff moved through a packing gland to a non contact level monitor. If routine inventory determinations are made, atypical losses can be identified and corrective action taken before large quantities of relatively expensive carbon are lost.

G. AMMONIA REMOVAL

Because of the corrosive nature of the brine used to regenerate the zeolite and the scale forming characteristics of the spent brine the design must facilitate scale removal operations and utilize corrosion resistant materials. If the steam stripping procedure is used to remove ammonia from the brine an acid cleaning system is required for the heat exchangers. The pressure drop across the heat exchangers should be continuously monitored and the service cycle terminated at an inlet pressure somewhat below the rated capacity to preclude gasket failures.

The interior of the stripping column must be accessible to facilitate inspection and cleaning. The fiberglass column at Rosemount was used in a total of approximately 400 stripping operations or approximately 1600 hr. After that period of operation the column was severely cracked, Figure 46, and the interior heavily coated with calcium carbonate scale. The white streaks visible in Figure 46a were related to leaks that developed near the top of the column.

The brine transfer equipment must be constructed of chemically resistant materials. They must also, however, be abraision resistant. The brine transfer

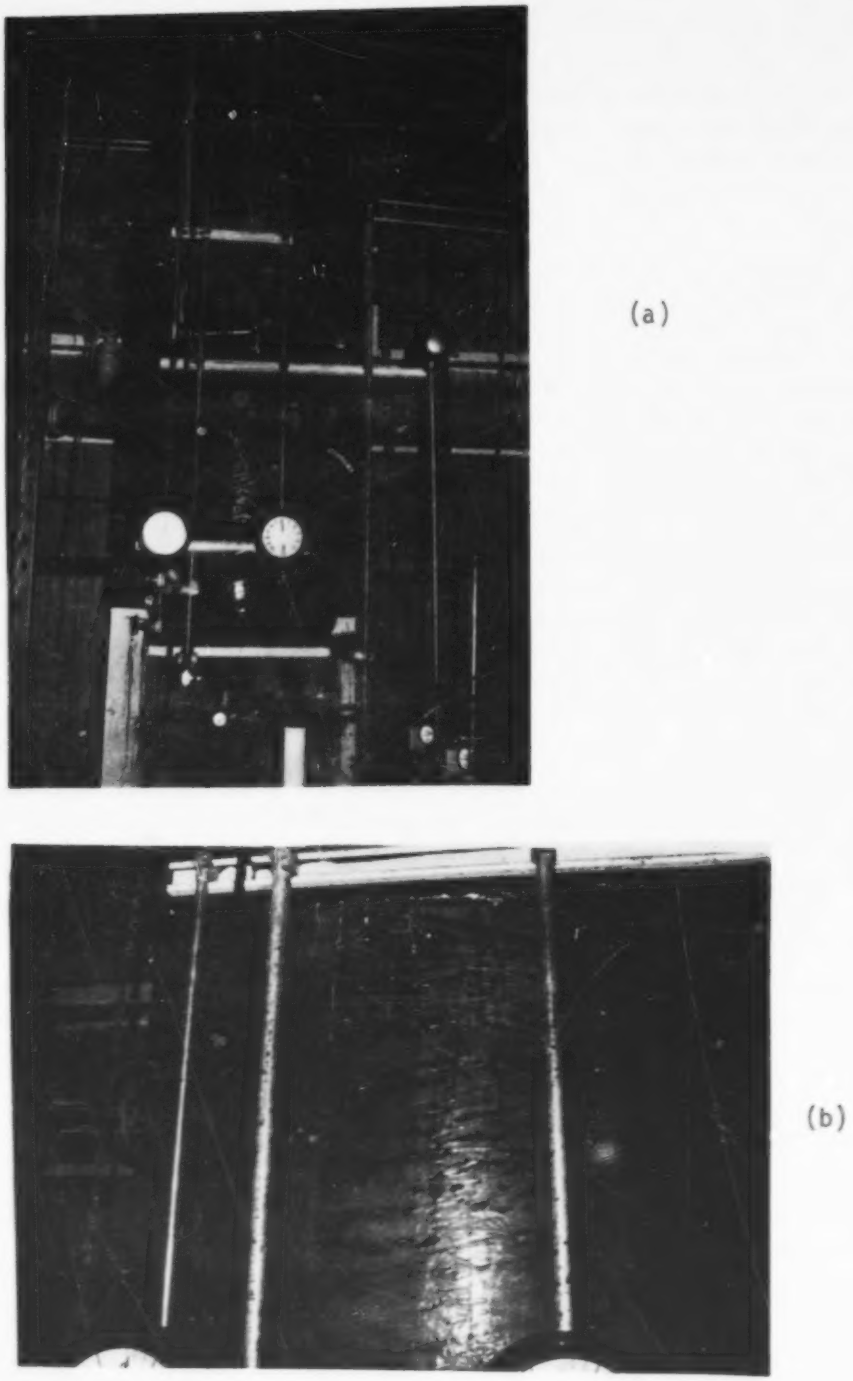


Figure 46. Cracks on outside surface of steam stripping column.

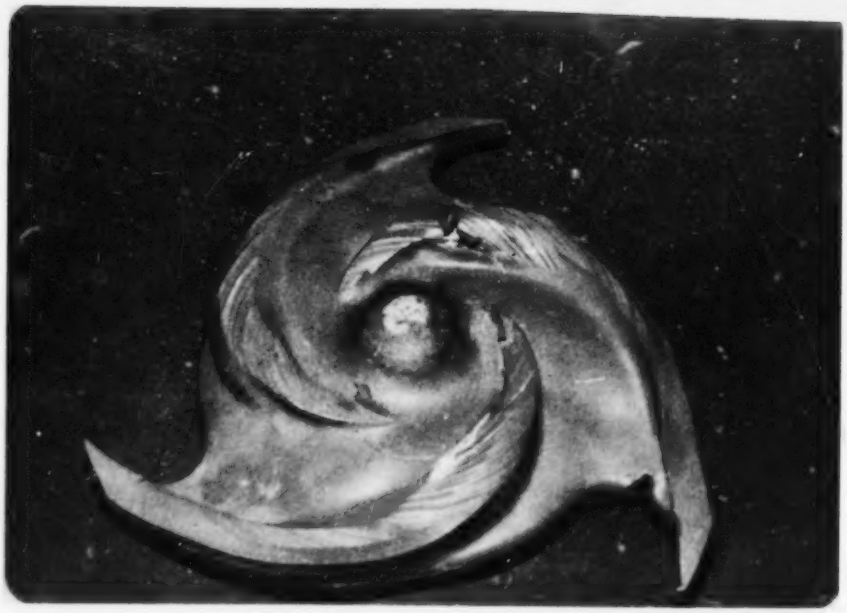
pump supplied by the contractor (Gould model 3196) was teflon lined ductile iron. The lining of the pump casing failed in July 1975 after only minimal service time and was replaced at a cost of $1,643. The lining on both the casing and the impeller failed in December 1976 - Figure 47. The pump was replaced with the identical model constructed of cast Gould-A-Loy 20 and no further problems were identified.

H. VALVES

Most of the process on/off valves at Rosemount are of the butterfly type. The valves in the brine area were supplied with stainless steel discs, all others were cast iron. Neoprene rubber was used for the seat material for all these valves. After approximately 1 year of operation many of the valves on the carbon column influent and effluent manifolds started to malfunction. When the valves were disassembled for repair it was observed that the rubber seats appeared to be slightly swollen. Because of this the valve operator did not have sufficient torque to open the valve once completely closed. All of the carbon column valves were repaired with Hi-Car rubber seats. During 1976 the same problems were encountered with the valves serving the ion exchange columns and the neoprene seats were again replaced with Hi-Car rubber. No problems have been encountered with the new seat material. Because of the nature of the physical-chemical processes it is not unusual to have several flow streams carried alternately in the same conduit. It is essential that the valves maintain drip tight integrity. If this is to be accomplished the question of material compatibility must be thoroughly addressed.

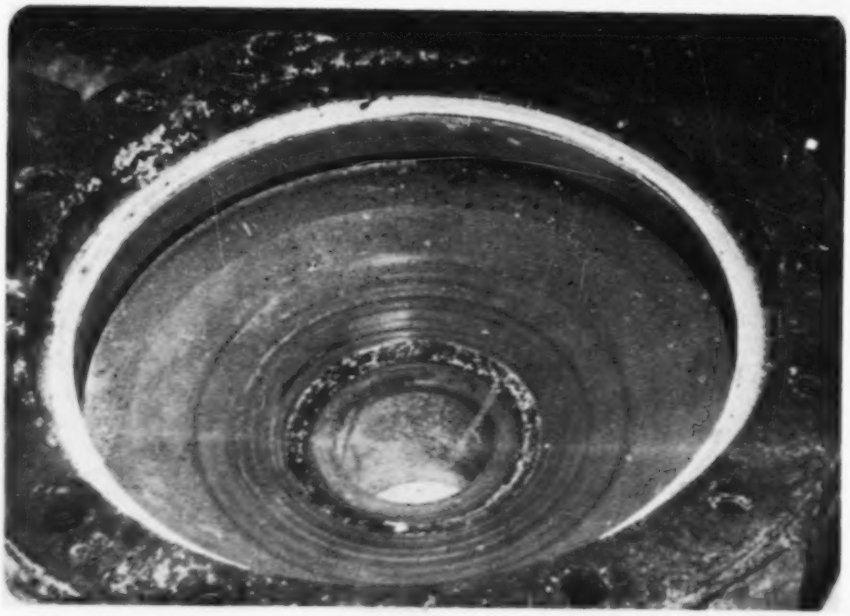
Because of the frequent backwashing and column sequencing procedures a treatment facility of this type will utilize a large number of remotely operated on/off valves. It is advisable that the limit switches, used to provide remote indication of valve position, be attached to the valve stem or shaft rather than to the operator. In the event that the coupling between the operator and the valve fails, this would insure that the remote indicator would provide a true indication of valve position.

In order to enhance plant reliability and the safety of the operating and maintenance staff it is essential that all valves be readily accessible. From an operating viewpoint it would be best to locate all valves within easy reach from the floor. However, in the event that some valves are located above structures (filters, carbon contractors, etc.), they must be accessible by means of properly protected ladders and catwalks.



(a)

Impeller



(b)

Casing

Figure 47. Teflon lined brine transfer pump.

XIV. OTHER OBSERVATIONS

A. DATA CORRELATIONS

Prior to and during the early months of the demonstration project it appeared that several parameters, which might be monitored on a continuous basis, could be used to indicate the performance level of the unit processes and the plant overall. Linear regression and correlation coefficients were calculated for parameters which appeared to be related or where good correlation was anticipated. These coefficients, based on two years of laboratory data are summarized in TABLE 64.

A number of parameter pairs which are often assumed to be related did not exhibit strong correlation. Several of the raw wastewater characteristics as well as the performance characteristics of the clarification and carbon adsorption processes were found to exhibit little dependence on the daily flow. Based on these calculations it does not appear that the treatment efficiencies of the processes used were significantly affected by loading in the ranges observed.

The results of the regression calculations were not entirely negative. They did indicate that both turbidity and TOC could be used to characterize the quality of several process streams. If reliable instruments are available both parameters could be used to monitor performance on a continuous basis. This might provide potential to improve process control.

B. PERFORMANCE MONITORING

1. General

The original funding agreement for the demonstration project specified that several parameters be monitored on a continuous basis. The hardware and sample streams are described in section V of this report. Although the instruments were not formally evaluated some considerable operating experience was obtained. The following paragraphs discuss the general performance characteristics of the hardware used and their applicability at the Rosemount AWTP and other facilities of the same type.

2. Temperature

No problems were encountered with the temperature monitor during approximately three years of operation. The total operating and maintenance cost for the system averaged approximately $200.00 per year including the cost of the chart paper, periodic calibration checks and semi-annual maintenance of the

TABLE 64. SUMMARY OF LINEAR REGRESSION CALCULATIONS

	X		Y			
Line	Sample	Parameter	Sample	Parameter	Line of Best Fit	Coefficient of Correlation
1	Raw	Q!	Raw	NH_3-N	Y = 48 - 0.05X	- 0.29
2	Raw	Q!	Raw	SS	Y = 281 - 0.18X	- 0.05
3	Raw	Q!	Raw	COD	Y = 622 - 0.47X	- 0.08
4	Raw	Q!	Raw	Total P	Y = 18.5 - 0.02X	- 0.30
5	Raw	Total P	Raw	Total SP	Y = 3.2 + 0.46X	0.72
6	Raw	TOC	Raw	BOD	Y = 34 + 1.71X	0.85
7	Raw	TOC	Raw	COD	Y = 65 + 4.21X	0.75
8	Raw	COD	Raw	SCOD	Y = 119 + 0.047X	0.30
9	Raw	COD	Raw	Total P	Y = 7.1 + 0.009X	0.59
10	Raw	COD	Raw	SS	Y = - 40 + 0.55X	0.84
11	Raw	COD	Raw	BOD	Y = 59 + 0.30X	0.84
12	Raw	SS	Raw	Total P	Y = 8.6 + 0.013X	0.55
13	Clar	Total P	Clar	Total SP	Y = - 0.08 + 0.53X	0.82
14	Raw	SS	Clar	SS	Y = 18 + 0.007X	0.08
15	Raw	COD	Clar	COD	Y = 86 + 0.01X	0.09
16	Raw	SCOD	Clar	SCOD	Y = 29 + 0.23X	0.42
17	Raw	Total P	Clar	Total P	Y = 0.9 + 0.002X	0.01
18	Raw	Total SP	Clar	Total SP	Y = 0.3 + 0.02X	0.13
19	Raw	SS	Clar	COD	Y = 90 + 0.002X	0.01
20	Raw	SS	Clar	SCOD	Y = 64 - 0.008X	- 0.06
21	Raw	SS	Clar	Total P	Y = 1 - 0.0002X	- 0.06
22	Raw	SS	Clar	Total SP	Y = 0.5 - 0.0002X	- 0.10

! daily flow in thousand gallons

TABLE 64 (Continued)

Line	X Sample	X Parameter	Y Sample	Y Parameter	Line of Best Fit	Coefficient of Correlation
23	Clar*	COD	Clar*	Turb	Y = 0.6 + 0.87X	0.20
24	Clar*	Total P	Clar*	Turb	Y = 3.2 + 8.1X	0.25
25	Clar*	SS	Clar*	Turb	Y = 2.9 + 0.33X	0.26
26	Clar*	SCOD	Clar*	Turb	Y = 1.5 + 0.11X	0.18
27	Clar**	COD	Clar**	Turb	Y = - 1.3 + 0.11X	0.47
28	Clar**	Total P	Clar**	Turb	Y = 4.0 + 4.2X	0.47
29	Clar**	SS	Clar**	Turb	Y = 4.8 + 0.22X	0.52
30	Clar**	SCOD	Clar**	Turb	Y = 6.1 + 0.05X	0.18
31	Clar	Q!	Clar	Turb	Y = 11.4 - 0.008X	- 0.06
32	Clar	Q!	Clar	SS	Y = 19 + 0.002	0.01
33	Clar	Q!	Clar	COD	Y = 91 + 0.000X	0.00
34	Clar	Q!	Clar	SCOD	Y = 59 + 0.013X	0.03
35	Clar	Q!	Clar	BOD	Y = 47 - 0.02X	- 0.05
36	Clar	Q!	Clar	SBOD	Y = 23 + 0.003X	0.01
37	Clar	Q!	Clar	TOC	Y = 23 + 0.004X	0.03
38	Clar	Q!	Clar	STOC	Y = 16 + 0.009X	0.06
39	Clar	Q!	Clar	Total P	Y = 1.2 - 0.0008X	- 0.07
40	Clar	Q!	Clar	Total SP	Y = 0.7 - 0.0009X	- 0.12
41	DMF_1	Turb	DMF_1	SS	Y = 5.6 + 0.59X	0.52
42	DMF_1	Turb	DMF_1	BOD	Y = 28 + 0.40X	0.18
43	Clar	SS	DMF	SS	Y = 5 + 0.21X	0.42

* Clarifier pH = 10.5

** Clarifier pH = 9.5

TABLE 64 (Continued)

Line	X		Y		Line of Best Fit	Coefficient of Correlation
	Sample	Parameter	Sample	Parameter		
44	DMF_1*	Turb	DMF_1*	SS	Y = 4 + 0.22X	0.68
45	DMF_1*	Turb	DMF_1*	BOD	Y = 17 + 0.02X	0.03
46	DMF_1**	Turb	DMF_1**	SS	Y = 3 + 1.03X	0.65
47	DMF_1**	Turb	DMF_1**	BOD	Y = 27 + 0.87X	0.29
48	GCC	SS	DMF_2	SS	Y = 3 + 0.31X	0.55
49	DMF_1	SBOD	GCC	SBOD	Y = 8 + 0.086X	0.20
50	DMF_1	BOD	GCC	BOD	Y = 10 + 0.28X	0.36
51	DMF_1	SCOD	GCC	SCOD	Y = 16 + 0.17X	0.31
52	DMF_1	COD	GCC	COD	Y = 23 + 0.19X	0.27
53	DMF_1	STOC	GCC	STOC	Y = 3 + 0.28X	0.48
54	DMF_1	TOC	GCC	TOC	Y = 4 + 0.30X	0.49
55	DMF_1	SS	GCC	SS	Y = 5 + 0.14X	0.20
56	Clar	Q!	GCC	BOD	Y = 28 - 0.03X	- 0.14
57	Clar	Q!	GCC	SBOD	Y = 16 - 0.015X	- 0.13
58	Clar	Q!	GCC	COD	Y = 54 - 0.05X	- 0.16
59	Clar	Q!	GCC	SCOD	Y = 30 - 0.01X	- 0.04
60	Clar	Q!	GCC	TOC	Y = 11 - 0.005X	- 0.05
61	Clar	Q!	GCC	STOC	Y = 8 - 0.00025X	0.00
62	Raw	Temp, ^{o}F	GCC	COD	Y = - 11 + 0.89X	0.26
63	Raw	Temp, ^{o}F	% RCOD in GCC		Y = 112 - 1.2X	- 0.26
64	Eff	Temp, ^{o}F	GCC	COD	Y = 12 + 0.43X	0.14
65	Eff	Temp, ^{o}F	% RCOD in GCC		Y = 92 - 0.76X	- 0.17
66	GCC	H_2S-S	GCC	Turb	Y = 6.3 + 0.4X	0.21

TABLE 64 (Continued)

Line	X Sample	X Parameter	Y Sample	Y Parameter	Line of Best Fit	Coefficient of Correlation
67	GCC	H_2S-S	GCC	BOD	Y = 19 + 0.14X	0.05
68	GCC	H_2S-S	GCC	COD	Y = 33 + 1.56X	0.38
69	Eff	Turb	Eff	TOC	Y = 10 + 0.17X	0.15
70	Eff	COD	Eff	BOD	Y = - 3 + 0.46X	0.83
71	Eff	SS	Eff	BOD	Y = 13 + 0.52X	0.19
72	Eff	SS	Eff	COD	Y = 36 + 0.9X	0.18
73	Eff	SS	Eff	TOC	Y = 11 + 0.15X	0.08
74	Eff	Turb	Eff	BOD	Y = 11 + 0.58X	0.33
75	Eff	Turb	Eff	COD	Y = 32 + 0.91X	0.29
76	Eff	Turb	Eff	SS	Y = 0.4 + 0.42X	0.66
77	Eff	TOC	Eff	BOD	Y = 2.7 + 1.15X	0.76
78	Eff	TOC	Eff	COD	Y = 15 + 2.21X	0.78
79	Eff	Temp, °F	Eff	COD	Y = 42 - 0.04X	- 0.01
80	Raw	Temp, °F	Eff	COD	Y = 14 + 0.47X	0.13
81	All	SS	All	COD	Y = 93 + 1.03X	0.79
82	All	SS	All	BOD	Y = 43 + 0.37X	0.71
83	All	SS	All	TOC	Y = 27 + 0.15X	0.62
84	All	Turb	All	SS	Y = 4.1 + 0.68X	0.43
85	All	Turb	All	COD	Y = 49 + 1.14X	0.23
86	All	Turb	All	BOD	Y = 19 + 0.68X	0.26
87	All	Turb	All	TOC	Y = 13 + 0.22X	0.16
88	All	TOC	All	BOD	Y = - 2 + 1.95X	0.95
89	All	TOC	All	COD	Y = - 9 + 4.8X	0.91

TABLE 64 (Continued)

Line	X		Y		Line of Best Fit	Coefficient of Correlation
	Sample	Parameter	Sample	Parameter		
90	All	TOC	All	SS	$Y = -26 + 2.6X$	0.62
91	All	COD	All	BOD	$Y = 8 + 0.37X$	0.94
92	All	SS	All	BOD	$Y = 47 + 0.33X$	0.68
93	All	COD	All	SS	$Y = -35 + 0.63X$	0.78

recorder by the supplier on a contract basis.

Temperature effects may be significant when heated flows are recycled to various points in the plant on a regular basis. If the data are to be used on a regular basis however, they must be transcribed by hand and daily or other time based averages calculated. Excursions from the average must also be characterized. Data stored on paper charts are not readily accessible. At standard recorder speeds approximately 60 feet of inked charts are generated each month from each recorder. Based on the experience at Rosemount and other treatment facilities these data are not fully utilized because of the excessive cost of transcription. If data are collected with continuous monitors with the intent that they be used on a routine basis a computer based data acquisition system would be appropriate.

3. pH

The pH monitors functioned for approximately three years without problem. The multipoint recorder was serviced by the supplier on a semi-annual basis at a cost of approximately $100 per year. The cost of chart paper averaged $70 per year. After an initial period of frequent calibration checks, during which time it was determined that the instruments were stable, the checks were made at 3 to 4 week intervals and required approximately 12 man hours per year. The calibration adjustments fell in the range of 0 to 1.9 pH units and averaged 0.17 pH units.

It appears that continuous pH monitoring of at least several process streams is essential. The data produced with strategically located pH monitors can be used to identify malfunctions which would allow unwanted streams of acid or base access to the process flow streams. However, if the data are to be used they must be readily accessible and the comments of the previous section apply.

4. Conductivity and Chloride

Both the specific conductance and the chloride ion concentration along with the pH of the effluent stream were used to identify equipment malfunctions and operating procedures which allowed portions of the brine, used for zeolite regeneration, to enter the effluent stream. The conductivity monitor required no maintenance other than monthly calibration. The chloride monitor was calibrated every 2 to 3 weeks and the flow cell was cleaned annually.

The calibration adjustment for the conductivity monitor fell in the range of 0 to 19% full scale (0-5000 umho) and averaged 3.2% full scale. The required adjustments for the chloride monitor were substantially greater but were difficult to quantify because the instrument's logarithmic output (two decade scale 10-100-1000) was recorded on a linear scale. Because the two instruments were used to identify the time at which brine entered the effluent flow stream calibration status was not critical.

Specific conductance can be used only as a semi quantitative measure of effluent brine contamination because of the relatively high specific conductance of the OH^- ion. If more than a semi quantitative measure of brine loss is required a chloride ion monitor can be used if calibrated on a daily basis.

5. Total Organic Carbon

The performance characteristics of the process TOC analyzer were not adequate to allow or justify the use as a routine monitoring device. The original analyzer was plagued by failures of all major components. The manufacturer supplied a new instrument in October of 1976; however, the original infrared analyzer was used to make the CO_2 measurements. Although the new instrument was superior to the original device the span drift was significant. During a six week period of continuous operation, in the 0-50 mg/l TOC range, the instrument was calibrated on a daily basis and it was found that, while the zero drift was $\leq$ 4 mg/l 95% of the time, the span drift was $\geq$ 4 mg/l 40% of the time. After several unsuccessful attempts were made to have the infrared analyzer repaired a new infrared analyzer was purchased and installed. Based on 2 weeks of operation since that time the average daily zero drift and span drift were 1.5 mg/l and 0 mg/l respectively. It thus appears that, for at least the short term, reliable TOC measurements can be made on a continuous basis.

The data presented in TABLE 64 and Figure 48 demonstrate that TOC analyses may be used in lieu of the more time consuming analyses of organics when time is critical. Although the data presented in Figure 48 were generated with a laboratory TOC analyzer most process analyzers utilize the same measurement principle and should yield similar correlations with BOD and COD. MWCC experience with process TOC analyzers at the Rosemount AWTP and another facility however indicate that their maintenance costs may be significant and their use should be justified in terms of process control and/or required monitoring. Based on the experience obtained during the demonstration period the use of a process TOC analyzer is not justified for the Rosemount and similar facilities.

6. Turbidity

The turbidimeter functioned adequately during the project however problems encountered in the sample delivery system kept the instrument out of service

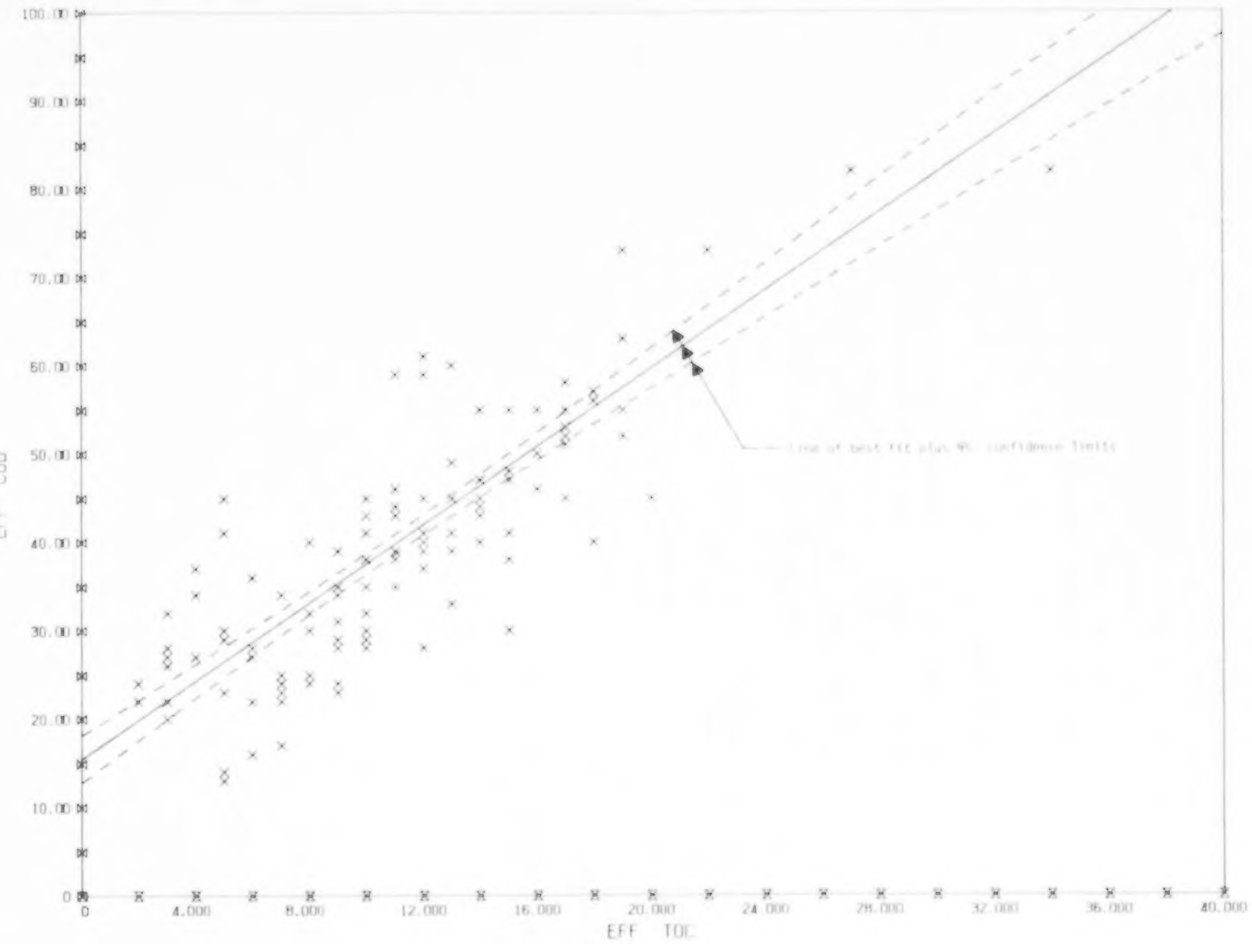


Figure 48. Relationship between COD and TOC for plant effluent.

for prolonged periods. The direct maintenance cost associated with the instrument during the demonstration was approximately $300 for the replacement of several circuit boards.

Laboratory turbidity measurements provided fair estimates of the suspended solids in several of the process streams (TABLE 64, lines 27, 29, 41, 44, 46, and 76). If the same relationships can be obtained with a flow through turbidimeter, use for monitoring process flow streams for estimates of suspended solids is justified. Because of the rapid response time of instruments of this type a number of sample streams can be monitored at 3 to 10 min. intervals as described in section V.C.

7. Ammonia and Phosphorus

Both the ammonia and phosphorus monitors functioned adequately; however, the instruments did require a significant maintenance effort. Because daily maintenance was required the monitors were generally operated only 5 days per week and the total operating time was approximately 6500 hrs. During that period the cost for consumables and repair parts was approximately $2500 and the labor requirement was approximately 550 man hours.

For a period of 100 operating days both monitors were calibrated on a daily basis. The zero and span drift for the ammonia monitor averaged approximately 1% and 10% of full scale respectively when operated in the range of 0-50 mg/l NH_3-N. The zero and span drift for the phosphorus monitor averaged approximately 3% and 6% of full scale respectively when operated in the 0 to 5 mg/l PO_4-P range.

Although the maintenance costs are substantial both ammonia and phosphorus monitoring may be justified when these compounds are being removed by physical-chemical means. Both the ammonia and phosphorus data could be used in process control schemes to ensure compliance with applicable standards and to minimize operating costs.

The equipment used for monitoring NH_3-N and P were standard laboratory automated chemistry modules (TABLE 9) and were not designed for continuous monitoring applications. At the time the above modules were purchased monitors designed for continuous service were available but their performance characteristics had not been sufficiently demonstrated and the combined cost of separate NH_3-N and P monitors was considerably higher than for the equipment purchased. It is possible that the performance characteristics and operating cost of monitoring equipment currently available are superior to those reported above.

REFERENCES

1. Dye, L.E., Feasibility Report on The Rosemount Advanced Wastewater Treatment Plant, Metropolitan Sewer Board, St. Paul, Minnesota, April 1972.
2. Process Design Manual for Carbon Adsorption, U.S. Environmental Protection Agency Technology Transfer, October 1973.
3. Standard Methods for the Examination of Water and Wastewater, 13th ed., American Public Health Association, New York, 1971.
4. Methods for Chemical Analysis of Water and Wastes, U.S.Environmental Protection Agency, Cincinnati, 1974.
5. Instruction Manual - Potassium Electrode, Orion Research, Inc., Cambridge, 1976.
6. Culp, R.L. and Culp, G.L., Advanced Wastewater Treatment. Van Norstrand Reinhold Co., New York, 1971.
7. Standard Method of Test for Ash in the Analysis Sample of Coal and Coke. 1975 Annual Book of ASTM Standards, Part 26, American Society for Testing of Materials, Philadelphia, 1975.
8. Standard Method of Test for Volatile Matter in the Analysis Sample of Coal and Coke, 1975 Annual Book of ASTM Standards, Part 26, American Society for Testing Materials, Philadelphia, 1975.
9. Viessman, W. Jr., Harbaugh, T.E. and Knapp, J.W., Introduction to Hydrology. Intext Educational Publishers, New York, 1972.
10. Dean, R.B. and Forsythe, S.L., Estimating the Reliability of Advanced Waste Treatment - Part 1. Water and Sewage Works, 123 (6):87-89, 1976.
11. Dean, R.B. and Forsythe, S.L., Estimateing the Reliability of Advanced Waste Treatment - Part 2. Water and Sewage Works 123 (7):57-60, 1976.
12. Beyer, W.H., Handbook of Tables for Probability and Statistics, 2nd ed. The Chemical Rubber Co., Cleveland, 1976.
13. Physical-Chemical Wastewater Treatment Plant Design. U.S. Environmental Protection Agency, Washington, D.C., 1974, 41 pp.

14. Tofflemire, T.J. and Hetling, L.J., Treatment of a Combined Wastewater by the low-lime process. Journal Water Pollution Control Federation 45(2), 210, 1973.

15. Parker, D.S., de la Fuente, E., Britt, L.O., Spealman, M.L., Stenquist, R.J., and Zadick, F.J., Lime Use in Wastewater Treatment: Design and Cost Data. EPA-600/2-75-038, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1975, 298 pp.

16. McClellan, T.J., Coagulation and Flocculation of Supra-Colloidal Particles Contained in a Tailings Basin Overflow. M.S. Thesis, Michigan Technological University, Houghton, Michigan, 1971.

17. Camp, T.R., Floc Volume Concentration. J. American Water Works Association 60:656, 1968.

18. Menar, A.B. and Jenkins, D., Calcium Phosphate Precipitation in Wastewater Treatment. EPA-R2-72-064, U.S. Environmental Protection Agency, Washington D.C., 1972. 96 pp.

19. Stamberg, J.B., Bishop, D.F., Warner, H.P., and Griggs, S.H., Lime Precipitation in Municipal Wastewaters. Federal Water Pollution Control Administration, Cincinnati, Ohio, 1969, 19 pp.

20. Wuhrmann, K., Objectives, Technology, and Results of Nitrogen and Phosphorus Removal Processes. In: Advances in Water Quality Improvement, E.F. Gloyna and W.W. Edkenfelder, eds. University of Texas Press, Austin, Texas, 1968 pp 21-48.

21. Kugelman, I.J. and Cohen, J.M., Physical-Chemical Processes. U.S. Environmental Protection Agency, Cincinnati, Ohio, 1972. 9 pp.

22. Stenquist, R.J. and Kaufman, W.J., Initial Mixing in Coagulation Processes. EPA-R2-72-053, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1972. 161 pp.

23. Vrale, L. and Jorden, R.M., Rapid Mixing in Water Treatment. Journal American Water Works Association 63(1) 52-58, 1971.

24. American National Standards Institute, Inc. American National Standard Acceptable Concentrations of Hydrogen Sulfide. ANSI Z37.2 - 1972, American National Standards Institute. New York, New York, 1972 8 pp.

25. Directo, L.S., Chen, C.L. and Kugelman, I.R. Pilot Plant Study of Physical-Chemical Treatment. Presented at Water Pollution Control Federation Conference, Denver, Colorado, 1974. 30 pp.

26. Weber, W.J., Hopkins, C.B. and Bloom, R., Physicochemical Treatment of Wastewater. Journal Water Pollution Control Federation 42(1): 83-99 1970.

27. Weber, W.J. The Role of Activated Carbon in Physicochemical Treatment. Presented at the US/USSR Symposium on Physical-Chemical Treatment of Wastewaters, Environmental Research Center, Cincinnati, Ohio, 1975, 49 pp.

28. Process Design Manual for Sulfide Control in Sanitary Sewerage Systems. U.S. Environmental Protection Agency Technology Transfer, October 1974.

29. Schimmel, C. and Griffin, D.B., Treatment and Disposal of Complex Industrial Wastes. EPA-600/2-76-123, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1976, 181 pp.

30. Westrick, J.J. and Cohen, J.M. Comparative Effects of Chemical Pretreatment on Carbon Adsorption. Presented at the Water Pollution Control Federation Conference, Atlanta, Georgia 1972, 37 pp.

31. Ford, D.L. Putting Activated Carbon in Perspective to 1983 Guidelines. Industrial Water Engineering 14(3):20-27, 1977.

32. Weber, W.J., Physiocochemical Processes for Water Quality Control. John Wiley and Sons, Inc., New York, New York, 1972, 640 pp.

33. Koon, J.H., and Kaufman, W.J. Ammonia Removal From Municipal Wastewaters By Ion Exchange. Journal Water Pollution Control Federation 47(3):448-465, 1975.

34. Optimization of Ammonia Removal By Ion Exchange Using Clinoptilolite. Water Pollution Control Research Series 17080 DAR 09/71, U.S. Environmental Protection Agency, Washington, D.C. 1971, 189 pp.

35. Hodgman, C.D., Weast, R.C., and Selby, S.M. Handbook of Chemistry and Physics, 39th ed. Chemical Rubber Publishing Co., Cleveland, Ohio, 1958, 3211 pp.

36. Ames, L.L. The Cation Sieve Properties of Clinoptilolite. American Mineralogist, 45:689-700, 1960.

37. Gardner, F.H., Williamson, A.R., and Skovronek, H.S. Naval Stores Wastewater Purification and Rinse by Activated Carbon. EPA-600/2-76-227, U.S. Environmental Protection Agency, Cincinnati, Ohio. 1976, 34 pp.

38. Von Dreusche, C. Jr. Process Aspects of Regeneration in a Multiple Hearth Furnace. Presented at 78th National Meeting of American Institute of Chemical Engineers, Salt Lake City, Utah, 1974.

39. Semmens, M.J. The Feasibility of Using Nitrifying Bacteria to Assist the Regeneration of Clinoptilolite. Presented at 32nd Purdue Industrial Waste Conference, West Lafayette, Indiana, 1977, 30 pp.